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Performance of simulated flexible integrated gasification polygeneration facilities. Part A: A technical-energetic assessment

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ABSTRACT

This article investigates technical possibilities and performances of flexible integrated gasification polygeneration (IG-PG) facilities equipped with CO_2 capture for the near future. These facilities can produce electricity during peak hours, while switching to the production of chemicals during off-peak hours.

Several simulations were performed to investigate the influence of substituting feedstock and production on IG-PG facility output, load and efficiency. These simulations were done using a detailed AspenPlus simulation model of a Shell entrained flow gasifier combined with conversion facilities. In this model carbon-rich feedstocks (oil residues, coal and biomass) were converted to a variety of products (H_2, electricity, FT-liquids, methanol and urea) using state-of-the-art technology. The size of the gasifier was limited to the equivalent of 2000 MW_{th} Il #6 coal input.

Overall efficiency of the simulated non-flexible configurations to convert pure coal or pure wood pellets to electricity ($40\%_{HHV}$ vs $38\%_{HHV}$), FT-liquids ($60\%_{HHV}$ vs $55\%_{HHV}$), methanol ($53\%_{HHV}$ vs $49\%_{HHV}$) or urea ($51\%_{HHV}$ vs $47\%_{HHV}$) are in good agreement with the literature. Using torrefied wood pellets instead of pure wood pellets reduces the penalty drop in efficiency compared to coal. Moreover, torrefied wood pellets have superior energetic density, handling and feeding compared to wood pellets.

In this analysis, the H_2:CO ratio of the sweet syngas was fixed to match FT-liquids criterion. As a result, overall CO_2 capture rates are low, around 56–65%, depending on the feedstock used. Still, especially with FT-liquids and methanol production, CO_2 emissions at the facility are significantly reduced; less than 20% of the carbon feedstock entering the facility is emitted with the flue gas. Applying biomass and CO_2 capture shows great opportunities to produce CO_2-neutral electricity or chemicals. When the biomass fraction exceeds 40% on an energy basis, production is CO_2-neutral, independent of what is produced.

Biomass can be co-fed up till 50% on an energy basis. Higher fractions cause significant fouling on cooling equipment. A small part-load penalty is observed during the substitution of coal by biomass. When changing from pure coal to pure wood pellets, the power case suffers a 2.5% efficiency drop, while all three chemical cases have an efficiency drop of less than 1%. At the same time total output is reduced to 67–69%, mainly because of the lower energy density of biomass. By over-dimensioning the gasifier and gas cleanup and optimisation section this drop can be eliminated.

The syngas can be tailored to the desired composition regardless of the used feedstock. Therefore, the chemical conversion sections only have to cope with a reduction in syngas flow and not with a change in syngas composition. Altering production between chemicals and electricity is possible, although the load of the conversion sections should remain between 40% and 100% to prevent operational problems. This gives a high degree of flexibility.

Complete substitution between chemical and power production while using the same feedstock is possible for the methanol and urea cases. The FT-liquids case is restricted to 60–100% load of the chemical conversion section to prevent that the gas turbine load is reduced below 40%.

The economic aspects of flexible IG-PG facilities are addressed in part B.

Abbreviations: AGR, acid gas removal; a.r., as received; ASU, air separation unit; CCS, carbon capture and storage; CGE, cold gas efficiency; d.a.f., dry and ash-free; EF, entrained flow; EP, Eucalyptus pellets; GT, gas turbine; HHV, higher heating value; HRSG, heat recovery steam generation; IGCC, integrated gasification combined cycle; IG-PG, integrated gasification polygeneration; IP, intermediate pressure (steam); LHV, lower heating value; LPMeOH^{TM}, liquid phase methanol synthesis process; LP, low pressure (steam); MP, medium pressure (steam); ppm, parts per million; PSA, pressure swing adsorption; RPS, rotating particle separator; SOTA, state-of-the-art; t, metric tonne; TOPS, torrefied (wood) pellets; TSA, temperature swing adsorption; WGS, water-gas shift; wt%, weight percent.

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Contents

1. Introduction

In 2008 global fossil fuel consumption was around 410 EJ. About 40% of this came from crude oil [1,2]. These fossil fuels power the economies of today. However, their use has several drawbacks. Fossil fuel resources are finite and their consumption results in large CO_2 emissions (unless carbon capture and storage is applied) as well as other environmental effects. Moreover, their consumption may result in high import dependence which may threaten supply security, for example in the European Union. Lastly, large-scale implementation of solar and wind power – so-called intermittent renewables – requires large-scale back-up electricity generation capacity. The output of back-up facilities must be easily adjustable in case solar and wind electricity production changes [3]. Also, the operating hours of these back-up facilities will be relatively low compared to base power plants.

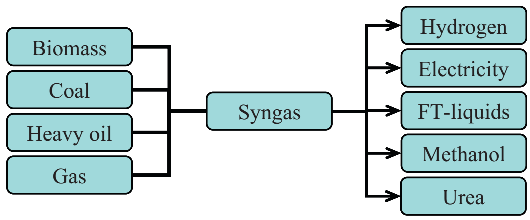


Fig. 1. Conversion of carbon rich feedstocks to different products.

A technology that can partly help to overcome these drawbacks is gasification. Through gasification (renewable) biomass, coal and oil residues can be converted into syngas, which can subsequently be converted into electricity, transportation fuels and other chemicals (Fig. 1). Natural gas can be converted into syngas using reforming.

To reduce CO_2 emissions, CO_2 can be extracted from process- and flue gases, compressed to 30–110 bar, transported and stored in underground fields (generally known as CCS). In this way, electricity and hydrogen can be produced with minimal CO_2 emissions. CCS at IG-PG facilities can be accomplished at reduced costs [4,5]. The reason for this is twofold. First, the chemical conversion processes already require removal of CO_2 from the syngas. Second, the higher partial pressure of the CO_2 makes pre-combustion capture using physical solvents possible. This lowers the energy requirements for the CO_2 capture and subsequent compression. By combining gasification and CCS it is possible to drastically reduce global CO_2 emissions without drastic changes to the current infrastructure. Furthermore, co-feeding biomass allows the production and use of liquid hydrocarbon transportation fuels with no net CO_2 emissions [6,7]. At large biomass fractions even net negative CO_2 emissions can be achieved.

An attractive feature of gasification is its potential flexibility. Especially the following two combinations of flexible integrated gasification polygeneration (IG-PG) facilities are interesting:

(1) Coal substitution by biomass can facilitate the use of biomass as feedstock without being dependent on a constant biomass supply.
(2) Chemical plants producing mid-load electricity can reduce the need of dedicated natural gas fired mid-load power plants. Mid-load plants are operated when electricity demand exceeds the base load. As intermittent renewables electricity generation capacity is ever-increasing, mid-load power plants will increasingly have to act as back-up power plants for these intermittent renewables. This may create opportunities for flexible large scale IG-PG facilities.

Multi-feedstock polygeneration (XtY)[1] system integration has seen a lot of attention in literature. A brief overview of studies highlighting results relevant for this study is given below.

1.1. General

Cummer and Brown [8] investigated the major ancillary components, i.e. pre-treatment and gas cleaning, in integrated gasification combined cycle (IGCC) facilities. They investigated whether components specifically designed for coal could also handle biomass. They also determined the stage of development of ancillary equipment for biomass fired IGCC. Their conclusion is that most of the ancillary equipment is not yet commercially available.

1.1.1. Electricity production

Electricity production using coal-fired IGCC combined with CCS has received quite some attention. A recent study by NETL indicates that, based on state-of-the-art (SOTA) Shell gasification technology, a 636 MW_e IGCC can achieve an overall efficiency of $41\%_{HHV}$, while emitting 750 kg CO_2/MW h. Equipped with CO_2 capture, plant efficiency would at present drop to $32\%_{HHV}$ and CO_2 emissions to 90 kg CO_2/MW h [9]. Similar values are presented by the IEA-GHG with $40\%_{HHV}$ efficiency for Shell IGCC without CO_2 capture and $32\%_{HHV}$ with CO_2 capture [10,11].

1.1.2. Fischer-Tropsch fuels production

Although there is some variation in final efficiency among the literature, reported XtL conversion efficiencies lay between $46\%_{HHV}$ and $56\%_{HHV}$. Boerrigter calculated an overall conversion efficiency of $56\%_{HHV}$ for a non-capture CtL facility [12]. Larson et al. calculated an overall conversion efficiency for a non-capture CtL facility of $49\%_{HHV}$ (33% FT + 17% power). Introducing CO_2 capture and compression lowers the power output by 3% points. Converting a mixture of switch grass and coal to FT-liquids with CO_2 capture could be performed with an overall conversion efficiency of $47\%_{HHV}$ [13]. This efficiency was confirmed by a study of Hamelinck et al. They calculated that biomass can be converted to FT-liquids with CO_2 capture at $46\%_{HHV}$ efficiency (33% FT + 13% power) [14].

1.1.3. Methanol production

A CtM study based on the liquid phase methanol (LPMeOH™) demonstration process at the Eastman Chemical Company in Kingsport was performed by Heydorn and Diamond [15] showing a CtM efficiency of $60\%_{HHV}$ – assuming a gasification efficiency of $80\%_{HHV}$ and a syngas energy to electricity efficiency of $40\%_{HHV}$. Two different CtM configurations were analysed by Larson and Tingjin [16]. They calculated an efficiency of $47\%_{HHV}$ for once-through and $63\%_{HHV}$ for syngas recycle. A BtM efficiency was calculated by Williams et al. at $61\%_{HHV}$ [17]. Hamelinck and Faaij calculated a BtM efficiency of $51\%_{HHV}$ using a fluidised bed system [18].

1.1.4. Urea production

Urea synthesis via gasification has received little interest in open literature. Neelis et al. found an energy consumption of 21 GJ_{LHV}/t urea, resulting in a CtU energy conversion efficiency of $44\%_{LHV}$ [19].

1.2. Objectives

All these studies do not investigate variation in feedstock or production during the operation of the facility. A recent study from IEA GHG has made a first assessment of the flexibility between hydrogen and electricity production, but only at two different production ratios [10]. To our knowledge that is the only study which investigated the possibilities of flexible IG-PG facilities and the impact of flexibility on overall efficiencies. This study aims to fill this gap. The objectives of this study are therefore:

- First, to identify which bottlenecks occur when altering feedstock or production and how these bottlenecks can be resolved.
- Second, to analyse system behaviour and changes in the overall efficiency and in the energy, mass and carbon balances when altering feedstock or production.
- Third, to analyse the effect of flexibility on the CO_2 balance and emissions and compare the balance and emissions with those from dedicated XtY facilities.

[1] Facilities or systems where feedstocks are gasified and converted to products are referred to as XtY systems. The X is often substituted if a specific feedstock is used; biomass (BtY), torrefied biomass (TtY) or coal (CtY). The Y is often substituted if a specific output is produced; electricity (XtP), FT-liquids (XtL), methanol (XtM) or urea (XtU).

Table 1
Feedstock parameters used in this study.

	Unit[a]	EP [27]	TOPS [27]	Il #6 coal [9]
Heating value	MJ_{HHV}/kg a.r.	17.29	20.51	27.14
Moisture	wt% a.r.	10.00	5.00	11.12
Volatile matter	wt% dry	86.60	75.9	44.5
Fixed carbon	wt% dry	12.83	24.1	44.59
Ash	wt% dry	0.50	1.34	10.91
C	wt% d.a.f.	49.77	54.63	80.50
H	wt% d.a.f.	5.80	5.67	5.68
O	wt% d.a.f.	44.20	39.45	8.70
N	wt% d.a.f.	0.14	0.22	1.58
S	wt% d.a.f.	0.03	0.02	3.17
Cl	wt% d.a.f.	0.06	0.01	0.37

[a] The composition is on as received (a.r.), moisture free (dry) or moisture and ash free (d.a.f.) basis.

In this study the focus is on the performance of SOTA commercial technologies relevant for XtY systems. To study these systems, an AspenPlus flowsheet process model of a flexible IG-PG facility was build to generate energy and mass balances for different IG-PG facilities. To allow accurate modelling, data on the technical bottlenecks as minimal load constraints and part-load behaviour of the various components were collected.

The structure of this article is as follows: Section 2 describes the commodities, including the different feedstock characteristics. In Section 3 the different components, including feedstock pre-treatment, syngas cleaning and syngas conversion, and their operating conditions are described. Section 4 gives the process model and plant configurations. In Section 5 the case studies are defined. Section 6 gives the simulation results, including overall performance and the effect of flexibility on overall IG-PG performance. In Section 7 a discussion of the results and used assumptions is given. Finally, Section 8 contains the conclusions.

In this study, units are in SI-units and heating values are in higher heating value (HHV), unless stated otherwise.

2. Commodities

2.1. Feedstocks

For this study three representative solid carbon-containing feedstocks were selected. Eucalyptus pellets (EP), torrefied wood pellets (TOPS) and Illinois #6 coal. Their properties are given in Table 1. Short rotation trees, of which Eucalyptus is one of them, is a potentially important biomass source [20–22]. The biomass is directly pelletised or torrefied and then pelletised as this drastically improves biomass properties, like heating value and moisture content (Table 2). This results, among others, to a more efficient transportation and handling of the biomass [6,21,23,24]. Illinois #6 coal is a Bituminous coal type often used in and commonly used as reference coal [9,16,25,26].

2.2. End-products

Syngas can be converted into a wide variety of end-products. The main end-products evaluated in this study are electricity, FT-liquids, methanol and urea. An important intermediate product is hydrogen.

2.3. By-products

Two important by-products of XtY facilities are slag and elemental sulphur.

Slag is an inert glass-like material consisting mainly of minerals. It is formed as the ash in the feedstock melts and is subsequently quenched. Slag is, among others, used as concrete additive or road surface coating compound [28]. The amount and quality of the slag depends on the used feedstock. Coal generally produces more slag than woody biomass.

Almost all sulphur in the feedstock is recovered as elemental sulphur. It is used in a wide variety of processes, including vulcanisation, bleaching and the production of fertilizers and viscose [29]. Coal generally contains more sulphur than biomass.

3. Process description

In this section a technical description and part-load performance of the different components used in a flexible IG-PG facility is presented. For all components only commercial available technologies were investigated. The key processes are displayed in Fig. 2.

3.1. Plant flexibility

A flexible IG-PG facility must be designed to cope with alterations in feedstock, syngas flow and heat streams. Among others, it must be able to adjust the H_2:CO ratio. Physical limitations restrict the flexibility of the conversion processes. For example, separators are usually unable to operate properly below 40% volume load

Table 2
Overview of chemical properties of different fuel types.

	Il. #6 coal [16] (bituminous)	Fresh wood [27]	Wood pellets [33]	TOPS [34]
Heating value (MJ_{HHV}/kg a.r.)	27	10–11	18–20	20–24
Mass density of bulk material (kg/m^3)	600	350	500–650	850
Energy density of bulk material (GJ/m^3)	16–17 [34]	3–4	9–13	17–21
Moisture content (wt%)	11	50	6–10	1–5
C/H ratio (wt)[a]	13–16	~8	~8	9–10 [24]
C/O ratio (wt)[b]	7–10	1	1	1–2 [24]
Flammability[c]	o	++	+++	++ [35]
Hydrophobicity	+	–	–	+

[a] Gasifying a feedstock with a higher C/H ratio results in more CO and less H_2.
[b] A lower C/O ratio results in a lower heating value (HHV) of the fuel. In a gasifier it also results in lower external O_2 consumption.
[c] The flammability of coal was set on average. Wood is easier to combust, but its high moisture content lowers flammability slightly. During pelletisation, this moisture is removed. TOPS has a low moisture content, but during the torrefaction process it lost most of its volatile compounds, thereby lowering its flammability.

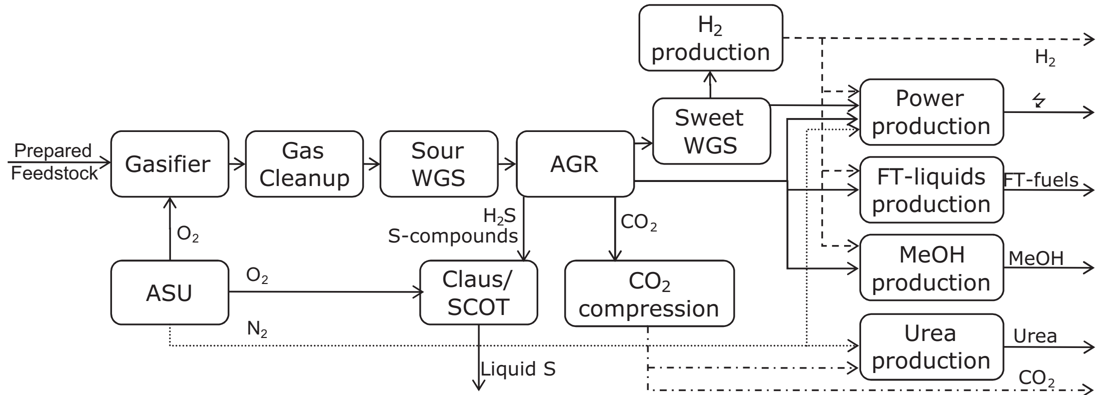


Fig. 2. Simplified process layout of a flexible IG-PG facility using SOTA commercially available technology. Waste, heat and recycle streams are not displayed.

[30,31]. Also, gas turbines have difficulties maintaining the desired flue gas exit temperature below this load [32]. In the subsequent paragraphs the load restrictions and part-load behaviour of the various components are discussed.

3.2. Pre-treatment

The gasifier has several minimal requirements concerning the feedstock (see Section 3.5). To fulfil these requirements, the feedstock, especially biomass, needs pre-treatment, like pelletising and torrefaction combined with pelletising. The effects of these techniques are given in Table 2. For comparison the average coal characteristics are also given.

The pre-treatment can be performed at the harvesting, central gathering or factory location [6,23]. It can also be divided over these locations. In this study it was assumed that the biomass is already torrefied or pelletised when arriving at the IG-PG facility. The most important pre-treatment techniques are described below.

3.2.1. Torrefaction

Torrefaction transforms biomass into a brittle, moisture-poor, hydrophobic solid with an increased energy density compared to the original biomass [23,36]. In short: the biomass becomes more coal-like. Another benefit is better storage properties as deterioration is much slower than raw biomass. The torrefaction process consists of four stages: heating, drying, torrefaction and cooling. During the torrefaction stage, the biomass is heated in a moving bed reactor to 200–300 °C for several minutes. H_2O, CO_2 and other volatile compounds evaporate, resulting in an increased C/O ratio of the solid product. Besides this solid, a low caloric gas is produced. On average, the biomass will lose 30% of its mass and 5–10% of its energy, depending on the biomass used and the original moisture content. The low caloric gas is combusted and used as heating source. Additional biomass is combusted to supply the remaining heat demand, resulting in an almost completely self-sufficient system [23,36]. Torrefaction is a batch process requiring storage when applied in a continuous process, such as an IG-PG facility. Storage also fits to part-load operation of the facility. The torrefaction process is operated at full load, regardless of the IG-PG facility load. When the storage bins are filled, the torrefaction process is shutdown. It is restarted when the feedstock in the storage bins comes below a certain threshold value.

3.2.2. Pelletising

Pelletising increases the energy density of biomass feedstocks. This, among others, reduces transportation costs. During pelletising the biomass is highly compressed. If the biomass is torrefied, pelletising is mandatory except if the torrefied product is used immediately. When using raw biomass, the pelletising process consists of four stages: drying and grinding, steam conditioning (to soften the fibres), pressurisation and cooling. To prevent pellet disintegration, glue is needed. When pelletising at 150 °C the lignin in the biomass – which softens above 100 °C – is used as glue [23]. The energy density of the biomass increases by about a factor of 2 (Table 2). Electricity consumption of pelletising is 90–160 MJ_e/t input [23].

Combining torrefaction with pelletising results in a factor 3 increase in bulk energy density compared to raw biomass (Table 2) and a halving of the electricity consumption of the pelletising process. It is best to pelletise the biomass directly after the torrefaction step. This reduces the pelletising process to pressurisation and cooling only. Like torrefaction, pelletising is a batch process. Therefore, the same part-load behaviour applies.

3.2.3. Sizing

Sizing reduces the particle size of solid feedstocks, making transporting and feeding the feedstock more efficient and easier. It also improves drying, torrefaction and pelletising. Shell EF gasifiers demand particles smaller than 0.1 mm for coal and smaller than 1 mm for biomass. Biomass particles can be larger as they are more reactive than coal [37]. Hard brittle feedstocks (TOPS, coal) are ground using crushers. More fibrous feedstocks (raw or pelletised biomass) are ground using a hammermill. Grinding biomass to 1 mm and coal to 0.1 mm requires an energy consumption of 0.01–0.02 kW_e/kW_{th} – grinding biomass to 0.1 mm would require 0.08 kW_e/kW_{th} [37]. Sizing has good part-load behaviour. By reducing the speed of the conveyer belt and the sizing equipment, part-load operation can be achieved without lower efficiencies.

3.2.4. Drying

The high moisture content in raw biomass makes it unsuitable for gasification – fresh wood has a typical moisture content of around 50 wt% [27]. Drying is preferred before pelletising or long distance transport. Active drying of biomass below 10 wt% moisture is not practical as drying efficiency drops dramatically below that value [38]. Heat demand depends among others on initial and final moisture content, particle size and hydrophobicity of the feedstock. This heat can be supplied by waste heat of other processes. Drying has good part-load behaviour. Just like sizing the speed of the conveyer belt and the amount of heat added to the feedstock can be reduced. In this study the coal was dried to 2.5 wt% using waste heat of the IG-PG facility. Biomass did not receive further drying as its moisture content was already below 10 wt%.

Table 3
Feedstock pressurisation characteristics [37].

		Il. #6 coal[a]	TOPS[a]	EP[a]
CO_2 consumption of	Nm^3 CO_2/t feedstock	3.7	0.2	0.2
which:	to gasifier	2.1	0.1	0.1
	to atmosphere	1.6	0.1	0.1
Electrical consumption	kWh/t feedstock	46.7	42.5	42.5

[a] Based on a lock-hopper system with pneumatic feeding for coal and a hydraulic piston system with screw feeding for biomass.

3.3. Pressurising and feeding

Previous research indicated that pressurised gasifiers have significant advantages over atmospheric gasifiers. The most important advantages are smaller downstream equipment and having a gas stream at elevated pressure [14]. As most chemical conversion processes and the gas turbine operate at higher pressure, the latter advantage significantly lowers compression requirements. However, using pressurised gasifiers also requires pressurised feeding. For solid feedstocks, the pressurising and feeding systems depend on the particle density and size. Coal is usually fed by a lock-hopper system with pneumatic feed. Fresh (or pelletised) biomass needs a different feeding system as it is too fibrous. This can lead to blockage of the lock-hopper feeding system. Due to the larger particle size of biomass, a hydraulic piston system with screw feeding can be used. The torrefaction process destroys the fibres, allowing TOPS to use both feeding systems. However, screw feeding is more efficient than pneumatic feeding [37]. As both systems are batch systems, they are equipped with pressurised storage bins, allowing homogeneous and continuous feeding. As pressurising agent, CO_2 was used. Energy consumption depends on the used process and is needed for CO_2 compression and the hydraulic piston feeder system. In this study separate feeding system for the biomass and coal were used. Part-load operation only affects the frequency at which the storage bins are refilled. The effect of feeding the different feedstocks is given in Table 3 [37].

3.4. Air separation unit (ASU)

The use of pure oxygen (95%) instead of air leads to a lower syngas volume and higher syngas energy density. There are three different air separation techniques available to extract oxygen from air: (vacuum) pressure swing adsorption, membrane separation, and cryogenic separation [39]. Large scale membrane separation shows high potential but is still in the demonstration phase. Of the other options, cryogenic separation is currently the most economical option when high purity (>95%) and throughput (>2.5 t/h) are required [40,41]. Energy savings can be accomplished by integrating the ASU with the gas turbine. An integrated ASU receives (part of) its air already partly compressed from the gas turbine. Although this increases efficiency, plant flexibility and reliability is reduced due to the dependence of the ASU on the gas turbine [30,42]. Also, other modern techniques, like adjustable rotor blades, decrease these integration gains significantly. Therefore, in this study a non-integrated ASU is used. Power consumption of the ASU is 0.20 kWh/kg O_2 produced [43]. The pure oxygen is compressed in several steps to 120% of gasifier pressure at an isentropic efficiency of 75% [44]. Between the compression steps the oxygen is cooled to 40 °C. Large scale cryogenic ASU have good part-load behaviour till 50% load. Specific energy consumption is almost constant between 85% and 100% load [30].

3.5. Gasifier

The gasifier converts the feedstocks into a gaseous mixture of mainly CO and H_2. There are several types of gasifiers available. They differ in operating temperature, feedstock state (slurry or dry), type of oxidant (air or oxygen) and gasifier configuration. As a high quality syngas is required, an oxygen-blown EF gasifier is used as they typically operate at high temperature (>1500 °C) and pressure (20–70 bar) [28]. The most used EF gasifiers are the Shell and GE (formerly Texaco) gasifiers [45]. Although both gasifiers are commercially available, only the Shell gasifier [46] has had extensive biomass co-feeding runs. Also, the Shell gasifier is the most fuel flexible of the two. This is mainly caused by the difference in dry and slurry feeding. Slurry for the slurry-fed GE gasifier contains roughly 40% water (excluding the moisture in the feedstock), which must be evaporated. Therefore the feedstock needs a high heating value. The feedstock of a dry-fed Shell gasifier contains much less water. This results in less water evaporation, making the Shell gasifier suitable for feedstock with a low heating value, like biomass [47]. Therefore in this study a Shell EF gasifier is used. Typical operating conditions are given in Table 4. Currently, gasification facilities using a dry-fed Shell EF gasifier have been designed to co-feed biomass up to 30% on an energy basis [46]. Up to 50% might be possible for these gasifier, but this requires considerable adjustment of various processes, resulting in a reduction of overall efficiency, as well as very clean woody biomass and high energy coal, like anthracite. The different make-up of the ash in biomass – especially its chloride content – and its lower energy value are the main reasons for this limitation.

Table 4
Feedstock boundaries.

	Unit	Shell [48]
Particle size (coal-biomass)	mm	<0.1–1
Typical water content [49]	wt%	~2
Fuel state	–	Solid
Ash fusion temperature	°C	<1500
Ash content	wt% dry	8–40
Sulphur content	wt% dry	<7
Heating value	MJ_{HHV}/kg dry	>17
Maximum pressure [50][a]	bar	40

[a] Shell is developing gasifiers that operate at pressures above 40 bar. These gasifiers have not yet used biomass.

The chosen gasifier is a 40 bar, oxygen-blown, slagging EF Shell gasifier with a syngas exit temperature of 1500 °C. To reach the required high temperature a sub-stoichiometrical amount of oxygen is added, which reacts with the fuel. In the gasifier all oxygen and almost all carbon (>99.5%) reacts. Therefore, the syngas composition can be described by the following equations [28]:

$$CO + H_2O \leftrightarrows CO_2 + H_2 \quad \Delta H^{\circ}_{298\,K} = -41.2\,kJ/mol \tag{1}$$

$$CH_4 + H_2O \leftrightarrows CO + 3H_2 \quad \Delta H^{\circ}_{298\,K} = +206\,kJ/mol \tag{2}$$

Besides these main components, the syngas contains impurities like methane, N- and S-compounds, unconverted carbon particles, and slag are present. The outer wall of the gasifier is cooled with water to preserve the gasifier and the slag layer. This cooling extracts 4% of the feedstocks HHV [51]. This energy is used to generate steam. When syngas is cooled, contaminants (e.g. halides, carbon particles) in the syngas become sticky between 900 and 1100 °C. When using pure biomass as feedstock this stickiness can occur at temperatures as low as 500 °C [51]. This prevents the use of high fractions of

Table 5
Maximum allowable concentrations of main syngas impurities.

	FT reactor [55]
Total S	<1 ppmV
Total halides	<10 ppbV
Total N	<1 ppmV
Dust (d_{max} = 10 μm)	Almost completely removed
Tar	T_{dew} < 200 °C

Table 6
Optimal syngas compositions.

Product	Variables	Opt. value
H_2	H_2/CO	∞
Electricity	H_2/CO	N.A.
FT-liquids	H_2/CO	0.6–2
Methanol	$H_2/(2^*CO+3^*CO_2)$	1.3–1.4
NH_3	H_2/N_2	3
Urea	NH_3/CO_2	2.95

biomass. To prevent fouling of cooling equipment, the syngas is quenched to 850 °C [52]. As the syngas is shifted downstream a water quench is used. This lowers the steam consumption of the downstream water-gas shift (WGS) reactor.

Two important parameters to evaluate gasifier performance are carbon conversion and cold gas efficiency (CGE).

$$\text{Carbon conversion efficiency}\quad(\%) = \left(1-\frac{\text{unconverted carbon}\quad(\text{kmol/h})}{\text{carbon in feedstock}\quad(\text{kmol/h})}\right)\times 100 \tag{3}$$

$$\text{Cold gas efficiency}\quad(\%)=\frac{\text{HHV}_{\text{syngas}}\quad(\text{MW}_{\text{th}})}{\text{HHV}_{\text{feedstock}}\quad(\text{MW}_{\text{th}})}\times 100 \tag{4}$$

The carbon conversion efficiency is a measure for feedstock conversion. It is the fraction of carbon converted to gaseous compounds like CO, CO_2, CH_4 and HCN. Shell gasifiers can reach carbon conversion efficiency > 99.7%. The remaining carbon ends up in the slag or fly ash. The CGE is a measure of gasifier energetic efficiency. The CGE depends on the used feedstock and its moisture content. Modern dry-fed EF gasifiers have a CGE ranging between 80% and 83% for coal [16,53,54]. The gasifier is run at constant space-time volume. This means that a constant volume of syngas is leaving the gasifier each second, regardless of the used feedstock. By over-dimensioning the gasifier, the chemical energy of the syngas exiting the gasifier can be kept constant, even when using large fractions of biomass. When in part-load, the high temperatures must still be reached. But the relative heat loss will be higher than the previously mentioned 4%. Therefore, a larger fraction of the feedstock must be combusted, resulting in a lower CGE. Also, a sufficient amount of slag must be produced to maintain the slag layer [31].

3.6. Gas cleanup and syngas composition optimising

In the gas cleanup and composition optimising section contaminants are removed and the H_2:CO ratio of the syngas is adjusted to the desired value. The extend of purification and the desired H_2:CO ratio depend on the application. Table 5 gives the contaminant restrictions for the most demanding application of the syngas conversion processes – FT-liquids production – used in this study.

A gas cleaning and composition optimising train is described below in the order in which the syngas passes through. A downside of this train is the considerable variation in syngas temperature. The required heat exchangers are not described.

3.6.1. Cyclone

A cyclone or rotating particle separator (RPS) uses centrifugal and Coriolis forces to remove solid and condensed particles [56]. An average cyclone removes 90% of particles greater than 5 μm, while experiencing a pressure drop of 0.01 bar. With identical gas residence times a RPS removes 90% of particles greater than 0.5 μm [57]. The separation efficiency can be increased by increasing the residence time in the cyclone [51,58]. Operational data revealed that a combination of cyclone, filter and wet scrubber removes almost all particles from the syngas [46,52]. In this study it is assumed that the cyclone removes all solid particles. The cyclone has no energy consumption. Part-load behaviour affects cyclones in two ways. A reduction in inlet particle concentration lowers cyclone removal efficiency. However, this is offset by the lower initial particle concentration. The result is a more or less constant outlet particle concentration. A reduction in gas velocity lowers cyclone removal efficiency by about 2.5%point per 10% load reduction [59].

3.6.2. Candle filter

A filter also removes particles from the syngas. By cooling the syngas to 300 °C before the filter, alkali particles condense and are also removed from the syngas. A commonly used filter is the candle filter due to its fail-safe in the event of a filter breach. Regeneration is performed by back-flashing with an inert gas. This is commonly N_2 but in a CCS facility CO_2 can be used. Part-load behaviour has no negative effect on filter behaviour. Electricity is needed for the regular back-flashing. This consumption is negligible and not taken into account in this study.

3.6.3. Wet scrubber

A wet scrubber removes the remaining alkali and – small – particles by spraying the syngas with water. The water is collected at the bottom of the scrubber and is purified and recycled. Due to the high pressure, the temperature of the water can be around 175 °C. Syngas will leave the scrubber saturated. During part-load operation, the water flow is reduced. This keeps the cleaning effectiveness constant. The reduced water flow lowers energy consumption in the subsequent purification and recycling steps. Electricity is needed for the water pumps. This consumption is negligible and not taken into account. Minimal part load operation is 30% [60].

3.6.4. Water-gas shift (WGS) reactor

Each syngas conversion process has an optimal syngas composition (Table 6). The WGS reactor adjusts the H_2:CO ratio of the syngas using reaction (1). There are three types of WGS reactors: sour, high temperature sweet and low temperature sweet. The sour WGS reactor needs S-compounds in the syngas as the catalyst needs to be sulphided, while the sweet WGS reactors can only operate in absence of S-compounds. Therefore sour WGS reactors are located upstream the acid gas removal (AGR), while sweet WGS reactors are located downstream the AGR. Sour WGS reactors use Co/Mo catalysts at temperatures between 300 and 500 °C [28]. Besides the WGS reaction, S-compounds are converted to H_2S and HCN to NH_3 [61]. High temperature sweet WGS reactors (300–500 °C) use Cu-promoted, Fe/Cr or Co/Mo catalysts [28,51]. At these temperatures, around 97% of the remaining CO is converted. Low temperature sweet WGS reactors (180–270 °C) use Cu/Zn–Al oxide based catalysts. These catalysts are very sensitive to sulphur poisoning and demand very low S-concentrations (<1 ppmV). To maximise hydrogen production additional steam is added. In this study a sour WGS located upstream the AGR and a sweet WGS located downstream the AGR were used. In sour WGS reactors part-load operation or very clean biomass can lower the sulphur flow resulting in reduced performance of the catalyst. Besides this, part-load operation of WGS reactors has no negative effect on performance.

Table 7
Overview of process conditions and performance of Rectisol [64].

Technology	Final conc. (ppm S)	Operating temp. (°C)	Product
Rectisol	0.1–1	−30 to −60	H_2S

Table 8
Standard conditions.

Temperature	°C	25
Pressure	bar	1.013
Air composition	mol% N_2	0.770
	mol% O_2	0.207
	mol% H_2O	0.0136
	mol% Ar	9.17×10^{-3}
	mol% CO_2	2.97×10^{-4}

3.6.5. AGR

During the acid gas removal (AGR) S-compounds and CO_2 are separately removed from the syngas. When operated at low temperatures (<100 °C), the AGR also removes almost all H_2O from the syngas. Currently several commercial techniques can be used to extract the acid gases [62]. In this study the sour syngas contains a high CO_2 partial pressure. Also, deep sulphur removal is needed as downstream catalysts are very sensitive to sulphur poisoning. This makes the Rectisol process the preferred AGR technique. Its characteristics are displayed in Table 7. The Rectisol process consists of 2 absorbers, two flash drums, heat exchangers, coolers and compressors. Besides the sweet syngas, a pure CO_2 stream and a sulphur rich stream is produced. During operation the solvent slowly degenerates. Therefore, methanol is constantly added. The AGR uses 137 MJ_e/t CO_2 and 90 MJ_{th}/t CO_2 of low pressure steam [12,63]. When operating at part-load, two options are possible. Either the AGR continues to run at full capacity. This will improve capture efficiency, but will also increase energy consumption per kg CO_2 captured. Or, the recirculation flow of the solvent is reduced. This way capture efficiency and energy consumption per CO_2 captured remain almost constant.

3.6.6. Claus plant

S-compounds are converted to elemental sulphur in the Claus process. Here, 1/3 of the H_2S is combusted to SO_2 (reaction (5)) and any remaining hydrocarbons and ammonia are also combusted. Next are three catalytic beds where the remaining H_2S reacts with SO_2 to form elemental sulphur (reaction (6)) [65].

$$2H_2S + 3O_2 \leftrightarrows 2SO_2 + 2H_2O \quad \Delta H^{\circ}_{298\,K} = -519\,kJ/molH_2S \qquad (5)$$

$$2H_2S + SO_2 \leftrightarrows 3S + 2H_2O \quad \Delta H^{\circ}_{298\,K} = -74\,kJ/molH_2S \qquad (6)$$

After each bed the elemental sulphur is removed from the gas stream. Each catalytic bed operates at a lower temperature then the previous bed to maximise elemental sulphur formation. However, thermodynamic equilibrium prevents complete conversion to elemental sulphur at acceptable reaction rates. With three well tuned beds, recovery rates of 96–97% can be achieved.

The modelled Claus is oxygen-fired. This reduces the minimal sulphur concentration in the gas stream from 25 mol% for air-fired to 10 mol% [28]. The oxygen consumption of the Claus plant is around 2% of the oxygen demand of a gasification plant using Il #6 coal. The Claus process uses 190 MJ_e/t sulphur [66]. Part-load operation or substituting coal by biomass can results in a too low sulphur concentration in the gas stream. If total flow drops too low, the burner flame can become unstable or the temperature in the burner too low. To prevent this it might be necessary to recycle sulphur or add natural gas [28].

3.6.7. SCOT plant

The Claus off gas contains too much sulphur to comply with Western air pollution regulations. Therefore, a Shell Claus Off gas Treatment (SCOT) plant is used to recycle the sulphur from the off gas back to the Claus burner. The Claus off gas is fed into a catalytic bed, where all sulphur is converted to H_2S using hydrogen at elevated temperatures. The hot gas stream is cooled and H_2S is removed from the gas using an amine adsorption process. The H_2S is recompressed and recycled to the Claus unit. In this way >99.8% sulphur recovery rates can be achieved. The SCOT uses 100 MJ_e/t sulphur [66]. The SCOT plant is less sensitive to part-load operation than the Claus process. Therefore, a SCOT plant will have no problem with part-load operation.

3.7. Syngas conversion

After the syngas is cleaned, it can be distributed to the H_2 PSA and the syngas conversion sections. By altering the syngas flows to the chemical conversion sections and the H_2 PSA, production can be altered.

3.7.1. H_2 extraction

To produce pure hydrogen (a part of) the sweet syngas is shifted in a low temperature WGS reactor to increase the hydrogen concentration. Next, the hydrogen is extracted from the gas stream. The extraction process can use temperature or pressure swing adsorption (TSA or PSA) or membrane separation. The current standard is PSA [29]. The syngas passes through sorbents at which only hydrogen hardly interacts. All other components in the syngas bind to the sorbents. To regenerate the sorbents, the pressure is reduced, releasing the bonded contaminants [29]. As this is a batch process, several PSA columns are operated in parallel. The pure hydrogen can be used for ammonia production, H_2 cracking, SCOT reducing agent, adjusting the syngas H_2/CO ratio or as a valuable end-product. Part-load operation will reduce the frequency of regenerating the PSA columns [67].

3.7.2. Electricity

Electricity is generated by combusting syngas and combustible waste streams in a gas turbine. The syngas is diluted first with N_2 and, if necessary, with steam to lower the energy density of the syngas to the desired value. Excess air is injected to provide the necessary oxygen. The addition of N_2 and air also lowers the flame temperature, reducing NO_x formation. After combustion, the hot flue gas is used for steam generation. In this study an Alstom GT26 turbine was used. This model turbine is from 1993 and has more than 2 million hours of experience, making it suitable for IGCC application. The GT26 was selected as it has superior part-load behaviour because it can reduce the air flow down to 40% (Fig. 3) [68].

The air compressor efficiency was kept constant. To compensate for part-load behaviour the isentropic efficiency of the expander was adjusted using the following formula:

$$\eta_{Isentropic} = 0.65 + 0.3577 \times x - 0.1757 \times x^2 \qquad (7)$$

where x = ratio mass flow flue gas compared to nominal load.

Calibration was done using data from Alstom [68,69]. Part-load behaviour was based on part-load behaviour of the combined cycle (Fig. 3), assuming that the steam cycle compensates a third of the efficiency loss of the gas turbine [32].

3.7.2.1. Heat recovery steam generation (HRSG). In an IG-PG facility there are several processes – including coolers, syngas conversion reactors and compressors – that give off excess heat. Similarly, there are several processes that demand heat. All these heat streams

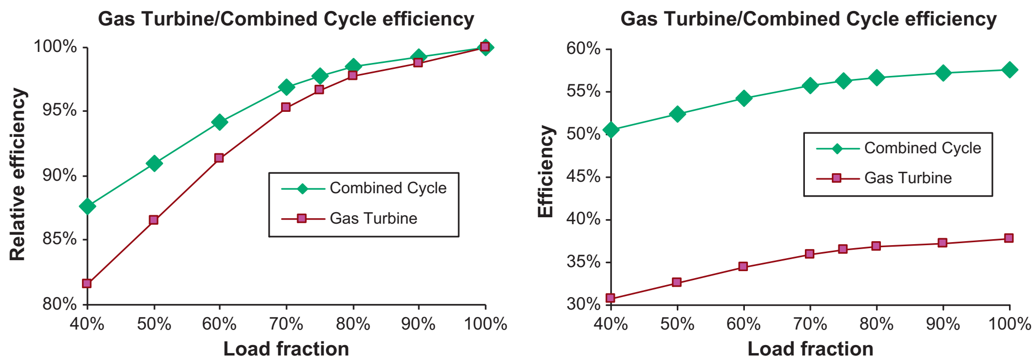


Fig. 3. Part-load efficiency gas turbine and combined cycle [32,68].

are combined in the heat recovery steam generation (HRSG) system. Any remaining excess heat is used for electricity generation by steam turbines [70]. The HRSG in this model uses three steam levels: low pressure (LP: 12.5 bar and maximal 368 °C), intermediate pressure (IP: 40 bar and maximal 550 °C) and high pressure (HP: 125 bar and maximal 550 °C). To improve the efficiency of the HRSG, the higher pressure steam is pre-heated with low temperature excess heat. Excess heat up to 200 °C is used to pre-heat internal used steam and LP steam. Excess heat between 200 °C and 450 °C is used to pre-heat IP and HP steam and to heat LP steam and internal used steam to the desired temperatures. Excess heat above 450 °C is used to generate IP steam. Only the heat from the first flue gas heat exchanger is used for HP steam generation. In part-load operation the water flows are reduced to maintain the desired temperatures of the steam.

3.7.3. *FT-liquids synthesis*

The FT-section converts syngas to linear alkanes and 1-alkenes. The process consists of an FT-reactor, heat exchangers, flash drums, distillation columns and H_2 cracker. The H_2 cracker is needed to upgrade the FT-liquids to transportation fuels that can be used in current car engines. During the FT-reaction about 20% of the chemical energy is converted into heat [14]. This heat is used for IP steam generation. The main FT-reaction (reaction (8)) is catalysed by Fe, Ru or Co [72]. The result is a hydrocarbon mixture with a composition that can roughly be described using a modified Anderson–Schulz–Flory distribution (Eqs. (9)–(11)) [14,71].

$$2nH_2 + nCO \rightarrow (\text{–}CH_2)_n^- + nH_2O \quad \Delta H = -165\,\text{kJ/mol} \tag{8}$$

$$S_{C5+} = 1.7 - 0.0024 \times T - 0.088\frac{[H_2]}{[CO]} + 0.18 \times ([H_2] + [CO]) + 0.0078 \times p \tag{9}$$

$$\alpha \approx 0.75 - 0.373 \times \sqrt{-\log(S_{C5+})} + 0.25 \times S_{C5+} \tag{10}$$

$$\alpha_{Cn} = \alpha^{n-1} \times (1 - \alpha) \tag{11}$$

where S_{C5+} = fraction hydrocarbons containing at least five carbons; T = temperature (K); $[H_2]$ and $[CO]$ = concentration of hydrogen and CO in the gas mixture; p = pressure of the gas mixture (bar); α = chain growth probability; α_{Cn} = fraction of hydrocarbons with n carbon atoms in FT-liquids mixture.

α increases with higher temperature, lower pressure, lower H_2/CO ratio, lower inert partial pressure or using Ru- or Co- instead of Fe-based catalyst. Maximum gasoline production is achieved at $\alpha = 0.76$ and maximum diesel production at $\alpha = 0.88$, but at these α large amounts of (too) light alkenes are produced (Fig. 4). Therefore, an $\alpha > 0.95$ is commonly chosen, followed by H_2 cracking of the wax to yield gasoline and diesel.

3.7.3.1. FT-liquids upgrading. The FT-synthesis yields the following products: gases (C_1–C_4), naphtha or gasoline (C_5–C_{11}), diesel (C_{12}–C_{20}), light wax (C_{21}–C_{30}) and heavy wax (C_{31+}) [73]. The gases are recycled to the gas turbine. To maximise economics – the desired products are gasoline and diesel – product upgrading is needed. This upgrading is similar to crude oil upgrading and FT-oil can be mixed with crude oil and used in existing refineries. The wax is led into a hydrocracker. Reaction rates of the hydrocracking depend on the size of the hydrocarbon as bigger hydrocarbons react faster than smaller ones. By using this principle, the alkane distribution can be tuned. The last step is distillation, resulting in naphtha (gasoline) and diesel [74]. The FT-liquids synthesis and upgrading has good part-load behaviour till 40% load, although small adjustments to temperature and pressure might be necessary [75].

3.7.4. *Methanol synthesis*

Current methanol production uses the Cu/Zn/Al catalyst promoted reaction of H_2 with CO and small amounts of CO_2 (reactions (12) and (13)) in a fixed bed or slurry reactor [65]. During the methanol reaction 15% of the chemical energy is converted into heat.

$$2H_2 + CO \leftrightarrows CH_3OH(l) \quad \Delta H^\circ_{298\,K} = -128\,\text{kJ/mol} \tag{12}$$

$$3H_2 + CO_2 \leftrightarrows CH_3OH(l) + H_2O(l) \quad \Delta H^\circ_{298\,K} = -131\,\text{kJ/mol} \tag{13}$$

A slurry reactor is used in this study as it has higher efficiencies – due to improved temperature control [15]. In the slurry reactor, the syngas diffuses from the gas phase through the liquid phase to the

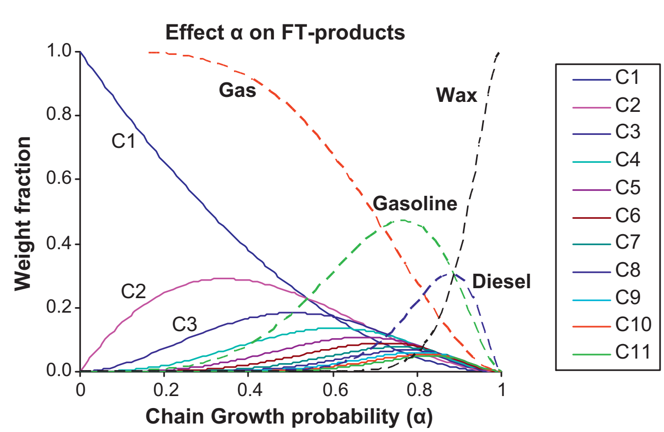


Fig. 4. Effect of α on FT-product distribution.

solid catalyst, where the reaction occurs. The primary function of the liquid is to regulate the temperature.

Purification of the methanol is done using a flash drum and three distillation columns. This leads to >99.5% pure methanol [74]. Unreacted reactants can be recycled back to the methanol reactor to obtain high overall conversion rates [61]. In this study all unreacted reactants were fed into the gas turbine. The methanol synthesis and upgrading has good part-load behaviour till 40% load [15].

3.7.5. NH_3 synthesis

Ammonia is synthesised by reacting H_2 and N_2 (reaction (14)) at high pressure and temperature (100–250 bar, 350–550 °C) using Fe-oxide catalysts. Due to thermodynamic limitations, high conversion rates can only be obtained by recycling the unused gas [61,65].

$$N_2 + 3H_2 \leftrightharpoons 2NH_3 \quad \Delta H^{\circ}_{298\,K} = -45.9\,\text{kJ/molNH}_3 \tag{14}$$

3.7.6. Urea synthesis

Ammonia is converted into urea by reacting it with CO_2 (reaction (15)) [77]. The CO_2 is supplied by the AGR and needs an additional compression step to reach the desired pressure.

$$2NH_3 + CO_2 \leftrightharpoons (H_2N)_2CO(s) + H_2O(l) \quad \Delta H^{\circ}_{298\,K} = -138\,\text{kJ/mol} \tag{15}$$

During the urea synthesis 26% of the chemical energy of the H_2 is converted into heat. The two most important urea synthesis processes are the Stamicarbon and the Snamprogetti process. Both processes use a two-step reaction. In this study the Stamicarbon process was used. It operates at 140 bar and 185 °C with a NH_3/CO_2 ratio of 2.95. This results in 60% CO_2 conversion (or 41% NH_3) [78]. High conversion rates are obtained by recycling the reactants. After decompressing the urea, its temperature must be kept below 130 °C to prevent decomposition of the urea. The ammonia and urea synthesis and purification has good part-load behaviour till 40% load. A potential problem is biuret formation. At part-load the biuret concentration increases. At high concentration – typical >1 wt% – the biuret needs to be removed as it is toxic to plants. The specific energy consumption will slightly increase when operating at part-load conditions [79].

4. Methodology: AspenPlus flowsheet model

4.1. Modelled components

The IG-PG facilities were modelled in the flowsheeting program AspenPlus. For each process there are – as indicated above – several alternatives. A short overview is given in Fig. 5. The processes used in the modelled flexible IG-PG facility are depicted in bold.

The selected processes were modelled using the parameters described in Table 9. An overview of the model lay-out is given in Fig. 6. The feedstock was gasified and converted to syngas. The syngas was cleaned and optimised (H_2:CO ratio). The sweet syngas was distributed over the different production sections. The offgas of the sweet WGS and FT-liquids, methanol and urea conversion sections were combusted in the gas turbine. Both FT-liquids and methanol sections were operated in a once-through mode. The urea section was operated in a recycle mode – both for the ammonia and urea synthesis. The pressure of the CO_2 for CCS is 110 bar, but the urea reactor operates at 140 bar. Therefore, the CO_2 for urea production underwent an additional compression step. To allow production flexibility, two distribution nodes distributed the gas flows. One node contained balanced syngas (H_2:CO ratio = 2.3) and was located between the guard bed and the conversion sections, bypassing the sweet WGS and H_2-PSA. The other node contained pure H_2 and was located between the H_2-PSA and the conversion sections. By altering the flow through these nodes the H_2:CO ratio was fine-tuned for the conversion processes. Especially the gas turbine behaved different at part-load. Therefore the isentropic efficiency of its expander was adjusted according to the load factor (Table 9). To maintain the desired pressure and temperature of the steam, the flow rates through the HRSG and steam turbines were adjusted. Regarding feedstock flexibility, the gasifier was the key determinant. Its gas generating capacity was kept constant for all cases.

The thermodynamic models used were an important aspect of the Aspen model to accurately predict the behaviour of the different components. Due to different process conditions, different thermodynamic models were used (Table 9). The standard conditions used in this study are given in Table 8.

4.2. Parameters

The process parameters of the modelled equipment were based on literature data, expert interviews and operational data of a commercially operated, Shell-gasifier based coal-fired power plant [80]. The parameters are given in Table 9.

5. Case studies

To determine how plant efficiencies and output change due to flexibility, several case studies have been simulated. The following limitations were enforced for all cases:

- Three types of feedstock are used: Illinois #6 coal, TOPS and EP. Three main coal:biomass fractions are used: (1) pure coal; (2) 50/50 energy% coal:biomass – the current technical limit; and (3) pure biomass.
- Gasifier capacity is kept constant at a gas generating capacity of 136 N m^3/s, equivalent to 2000 MW_{th} Illinois #6 coal input.[2]
- Syngas conversion processes offgas is directed to the gas turbine to generate electricity.
- The syngas H_2:CO ratio after the AGR is kept constant at 2.3, regardless of the used feedstock. This is done by adjusting the by-pass ratio of the sour WGS reactor.
- H_2 production in the sweet WGS is kept to a minimum.
- The energy density of the syngas entering the gas turbine is kept constant by dilution with N_2 and, if necessary, steam.
- All cases capture and compress CO_2, regardless of feedstock or production.
- Reactors and separators have – as described above – a minimal load demand of 40%. The gas turbine is calibrated till 40% load.

5.1.1. Reference cases

In this case study all components were optimally sized for the specific feedstock. There is no feedstock or production flexibility. This gives a reference point for the flexibility cases. The following simulations were performed:

o XtP: maximal electricity production;
o XtL: maximal FT-liquids production;
o XtM: maximal methanol production;
o XtU: maximal urea production.

[2] Previous research indicated that gasification facilities under 1000 MW_{th} suffer greatly from reduced economies of scale [12]. To ensure that both the chemical conversion and the power sections are sufficiently sized, a base size of 2000 MW_{th} coal input was taken.

Table 9
Overview AspenPlus model parameters.[a]

Pre-treatment and gasifier	
Drying	The fuel is dried to 10, 5 and 2 wt% moisture for EP, TOPS and coal respectively using excess heat from the flue gas.[b]
Milling	Grinding has an electric consumption of 0.015 kW_e/kW_{th}.[c]
Pressurising and feeding	2.1 and 0.1 N m^3 CO_2 needed to raise pressure from atmospheric to 48 bar at the inlet of the gasifier and feed the coal and biomass into the gasifier.[d]
ASU	Separated air into pure N_2 (100 mol% N_2) and oxygen (95 mol% O_2, 3.65 mol% argon and 1.35 mol% N_2). Electric consumption is 0.20 kWh/kg O_2.[e]
EF gasifier	RGibbs reactor. T = 1500 °C, p_{out} = 40 bar. Thermodynamic equilibrium. Carbon conversion = 99.5%. 4% HHV input heat loss. A moderator (H_2O; 42 bar, 246 °C) is added only with the pure coal cases.[f]
Wet quench	Mixer. T_{out} = 850 °C, $\Delta p = -2$ bar.[g]
Gas cleaning	
Cyclone	Removes all unconverted carbon particles. $\Delta p = -0.01$ bar.
Syngas cooler	Cools syngas from 850 °C to 230 °C. $\Delta p = -1$ bar.
Candle filter	Removes all carbon. $\Delta p = -0.5$ bar.
Wet scrubber	8 Stage column. Syngas fed from the bottom. Pressurised water fed from the top. Water flow adjusted to remove 95% of HCl. $\Delta p = -1$ bar.[h]
Sour WGS	Isotherm RGibbs reactor. T = 250 °C, final H_2:CO ratio = 2.3. $\Delta p = -1.5$ bar. H_2O:CO ratio in shift > 1.65.[i]
AGR	See AGR section below.
Guard bed	Removes any remaining H_2S and SO_2. T = 190 °C.
Sweet WGS	RGibbs reactor. T = 185 °C, final H_2:CO ratio = 100. $\Delta p = -1.5$ bar.
H_2-PSA	Separates 90% of the H_2, $p_{exit\ H2}$ = 24.9 bar, $p_{exit\ offgas}$ = 0.7 bar.[j]
AGR	
Rectisol process	A simplified version of the Rectisol process. Separation occurs based on literature references. $\Delta p = -6.6$ bar. 99.9% of the CO_2 is captured. Electric consumption is 137 MJ_e/t CO_2. Heat demand is 90 MJ_{th}/t CO_2.[k]
Claus burner	Combusts 1/3 of H_2S to SO_2. All other combustible components are combusted. T = 1300 °C. Gas is cooled to 330 °C.[l]
Claus conversion	3 Series of catalytic bed, cooler and gas/liquid separators. Catalytic beds are at 330 °C, 240 °C and 200 °C. $\Delta p = -1$ bar for each series. Electric consumption is 190 MJ_e/t S.[l]
SCOT	Catalytic reactor, separator and compressor. SO_2 is hydrogenated to H_2S. H_2S is separated, repressurised and mixed with Claus burner feed. Electric consumption is 100 MJ_e/t S.[l]
Power production[m]	
Syngas dilution	Pressurised N_2 from the ASU is added to lower syngas LHV to 4.3 MJ/N m^3. If N_2 supply is insufficient, pressurised steam is added.
Gas turbine	Adiabatic combustion at 24.9 bar. Turbine inlet temperature (TIT) = 1375 °C. Flue gas expanded to 1.5 bar. Isentropic efficiency expander = $0.65 + 0.3577 \times x - 0.1757 \times x^2$, where x is the mass ratio of the flue gas compared to design specification. At full load isentropic efficiency is 83.2%. Oxygen excess is 10%.[n]
HRSG	Flue gas is cooled in 3 steps to 150 °C. Remaining heat used as drying agent. A minimal temperature difference of 10 °C is maintained at each heat exchanger.
Steam turbines	Steam conditions are: HP = 125 bar, 550 °C; IP = 42 bar, 550 °C; LP = 12.5 bar, 368 °C. Isentropic efficiencies: HP = 85%, IP = 93% and LP = 89%. Condenser at 15 °C and 0.025 bar.[m]
Condenser	Condenses the steam to liquid water. The released heat is vented T_{exit} = 15 °C.[m]
FT-liquids production	
FT-reactor	T_{exit} = 230 °C, p_{exit} = 59 bar. Uses modified Anderson–Schulz–Flory distribution. $CO_{conv.}$ = 90%, offgas to power.[o]
FT flash	T_{exit} = 5 °C. p_{exit} = 54.6 bar.
FT decanter	T_{exit} = 55 °C. $\Delta p = -1$ bar.
FT cracker	T_{exit} = 250 °C. $\Delta p = -1$ bar, 10% stoichiometric excess H_2.

Table 9 (*Continued*)

Methanol production[p]	
Compressor	3 Stages. p_{exit} = 75 bar. Intercooled to 150–250 °C. Inter-stage Δp = 2% of pressure.
MeOH reactor	T_{exit} = 265 °C. p_{exit} = 69 bar, $T_{thermodynamic\ eq.}$ = 250 °C.
MeOH flash 1	T_{exit} = 40 °C. p_{exit} = 35 bar. Offgas to power.
MeOH distiller	10 Stage distiller. p_{exit} = 1 bar. Reflux ratio = 10.
MeOH flash 2	T_{exit} = 84 °C. p_{exit} = 3 bar.
Urea production[q]	
Compressors	1 Stage compressor. p_{exit} = 138 bar.
NH_3 reactor	T_{exit} = 300 °C. p_{exit} = 100 bar. Thermodynamic equilibrium.
NH_3 separator	4 Stage for H_2 and 2 stage for N_2. p_{exit} = 100 bar. Intercooled to 150 °C. Inter-stage Δp is = 2% of pressure.
NH_3 recycle	90% is recycled. 10% is bled.
Urea reactor	T_{exit} = 185 °C. p_{exit} = 140 bar. $CO_{2\ conv.}$ = 60%.[r]
Auxiliary equipment	
Compressors	$Eff_{isentropic}$ = 0.83. For oxygen this is 0.74. For CO_2 this is 0.8. $Eff_{mechanic}$ = 1.[s]
Water pumps	Eff_{pump} = 92.5%. $Eff_{mechanic}$ = 97%

[a] The used thermodynamic models were: solids for pre-treatment; PR-BM for gasifier, gas cleaning optimising, methanol section and power section; SR-Polar for AGR; SRK for FT-section; RK-Soave for urea section; and STMNBS2 for the steam section. Standard conditions are 25 °C and 1.013 bar.

[b] During pelletising the moisture content of the pellets is reduced to 6–9 wt%. Assuming that the pellets re-adsorb a little moist, it was expected that EP arrive with a moisture content of 10 wt%. Their hydrophilic nature prevents further – efficient – drying. TOPS have an initial moisture content of 5 wt%. Due to their coal-like hydrophobic nature, they are dried to 2 wt%. Coal is dried to 2 wt% [23,38].

[c] Coal particles were ground to 0.1 mm, biomass particles were ground to 1.0 mm. The bigger particle size, but more energy intensive grinding of biomass results in more or less similar energy consumption per kW_{th} compared to coal [37].

[d] The values were based on the assumption that particles with a size of 0.1 mm for coal and 1.0 mm for TOPS and biomass are used. An additional 1.6 and 0.1 N m^3 CO_2 was eventually vented to prevent syngas from escaping [62].

[e] A standard non-integrated cryogenic ASU was simulated. A conservative energy consumption of 0.2 kWh/kg O_2 was used [43].

[f] 4% of feedstock HHV is removed due to cooling. This is on the high side for Shell EF gasifiers. The high temperature justifies thermodynamic equilibrium. Literature data and expert interview indicates Teq. = 1500 °C should be used [81]. The amount of moderator was varied to maximise overall output in the XtL and XtM cases. Only if pure coal was used, a moderator was needed. Even with 20% biomass, adding a moderator reduced overall output.

[g] Standard quench temperature is 900 °C. As biomass increases fouling problems, 850 °C was used [52].

[h] The wet scrubber conditions were based on data from Chiesa et al. [25].

[i] The H_2:CO ratio was based on the FT process as it requires the largest CO fraction. The ratio will be obtained after the AGR. The H_2O:CO ratio was set at 1.65 instead of the conventional 2.0 based on expert interview [82].

[j] It was assumed that the pure H_2 has no pressure drop. The H_2-PSA behaviour was based on Shell data [83].

[k] The Rectisol process conditions were based on data from Kreutz et al. [5].

[l] Based on the theoretical optimal performance of a Claus unit. The high temperature was needed for NH_3 combustion. The beds operate at thermodynamic equilibrium [66].

[m] Based on open literature, NUON experimental data and expert interview [14,80–82]. Pressure losses in the steam cycle due to part-load operation were ignored.

[n] Based on Alstom data and expert interview at KEMA [32,68].

[o] FT-conversion based on previous literature. The Anderson–Schulz–Flory distribution was adjusted for increased C_1–C_4 production [14].

[p] Methanol conversion based on data from Heydorn and Diamond [15].

[q] The urea section was modelled based on AspenTech corporate flowsheets [84].

[r] The CO_2 needed for the urea synthesis was first compressed to 110 bar together with the CO_2 for CCS. It was subsequently compressed to 140 bar required for the urea synthesis.

[s] Different compressor characteristics were used. To compress air for the gas turbine a 4 stage compressor with an exit pressure of 101% of gas turbine pressure and no intercooling was simulated. To compress syngas after the cleaning and optimising section a 2 stage compressor with an intercooler to 100 °C and a 2% pressure drop was simulated. To compress CO_2 to 110 bar a 5 stage compressor with intercoolers to 40 °C was simulated.

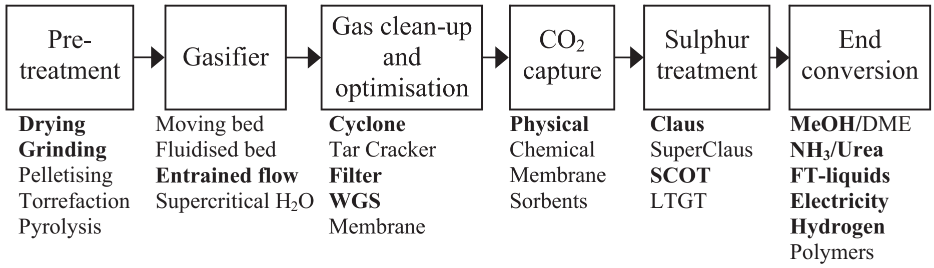


Fig. 5. Types of processes used in gasification.

5.1.2. *Effect of feedstock substitution*

In this case study coal was gradually replaced by either Eucalyptus pellets (EP) or torrefied wood pellets (TOPS). As more coal is being replaced by biomass, the total chemical energy of the syngas drops, i.e. the syngas energy density decreases. This affects chemical and electricity output. The following simulations were performed:

- o 1-P: maximal electricity production while increasing the biomass fraction;
- o 1-L: maximal FT-liquids production while increasing the biomass fraction;
- o 1-M: maximal methanol production while increasing the biomass fraction;
- o 1-U: maximal urea production while increasing the biomass fraction.

5.1.3. *Effect of end-product substitution*

In this case study the chemical output was reduced in favour of electricity output by diverting syngas form the chemical production section to the power section. The feedstock remains identical. The following simulations were investigated:

- o 2-L: switching between electricity and FT-liquid production;
- o 2-M: switching between electricity and methanol production;
- o 2-U: switching between electricity and urea production.

5.1.4. *Effect of changing production by feedstock substitution*

In this case study biomass was replaced by coal to increase electricity output. It was assumed that the base case was biomass conversion to chemicals. This can occur when biomass is less expensive than coal, by e.g. biomass subsidy or CO_2 tax. When electricity prices rise, biomass was replaced by coal. The additional energy in the syngas was directed to an over-dimensioned gas turbine, resulting in increased electricity output. The chemical production was kept at a constant level. This way the chemical production section always operated at optimal performance and capacity. The following simulations were investigated:

- o 3-L: producing FT-liquids at a constant level and increasing electricity production by substituting biomass with coal;
- o 3-M: producing methanol at a constant level and increasing electricity production by substituting biomass with coal;

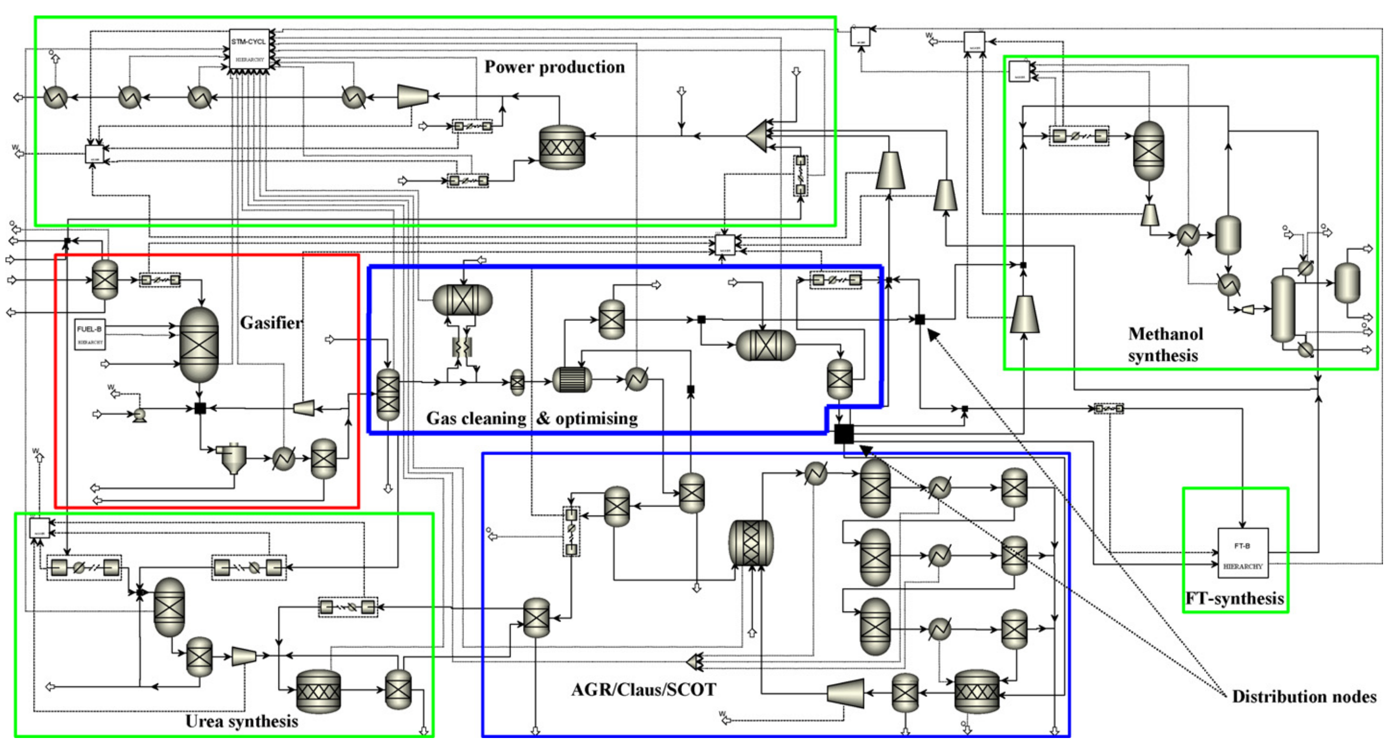


Fig. 6. Schematic overview of the AspenPlus flowsheet of the IG-PG facility. The solid lines are mass flows, the dashed lines are heat flows and the striped lines are electricity flows.

Table 10
General parameters gasifier.

	Unit	EP	TOPS	Coal
Feedstock input	MW_{th} fuel	1464	1696	2000
	t dry fuel/h	274	283	236
Oxygen consumption	kg O_2/kg dry fuel	0.61	0.59	0.88
	kg O_2/GJ_{th} fuel	32	28	29
CGE	%	70	76	82
Raw syngas composition exiting gasifier				
H_2O	(mol% wet)	25	14	1
H_2	(mol% dry)	27	29	30
CO	(mol% dry)	54	61	66
CO_2	(mol% dry)	17	9	0
Other	(mol% dry)	2	2	3

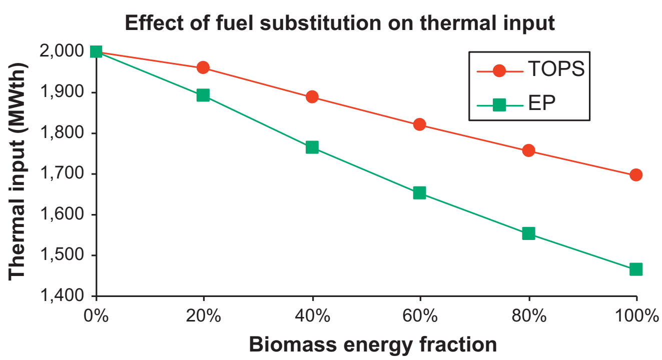


Fig. 7. Effect biomass substitution on thermal input while maintaining constant gas volume generation in the gasifier.

o 3-U: producing urea at a constant level and increasing electricity production by substituting biomass with coal.

6. Results

The assumption that the gas volume flow of the gasifier is the limiting factor has a direct impact on the input of the gasifier and thus the gas flows throughout the entire IG-PG facility.[3] The differences between feeding biomass or coal are displayed in Table 10.

Total chemical energy input drops with increasing coal:biomass fraction. Compared to pure coal the energy drops to 85% for pure TOPS and 73% for pure EP. Thermal input at different coal:biomass fractions is displayed in Fig. 7. As a moderator is added only with the pure coal case, the thermal input at pure coal is lower then expected when looking at the thermal input of the other coal:biomass fractions.

The feedstock mass input in the gasifier increased to 120% for pure TOPS and 116% fore pure EP compared to pure coal.

However, as the specific oxygen consumption of the biomass is much lower (0.6 t O_2/t dry biomass) than that of coal (0.9 t O_2/t dry coal), absolute oxygen consumption is reduced to 81% when biomass is used instead of coal. The lower specific oxygen consumption is a result of the higher oxygen-atom fraction in the biomass.

The lower energy density of biomass results in a higher fraction of the feedstock that is combusted. This leads to higher CO_2 and H_2O concentrations in the raw syngas and lower cold gas efficiencies (CGE) of the gasifier.

Another difference between biomass and coal is ash composition and input. Slag (molten ash) production drops by a factor 7 with pure TOPS and a factor 19 with pure EP compared to pure coal. When slag production is too low, the protective slag layer at the inside of the gasifier may disappear. This can damage the gasifier and reduce availability. Also, the ash composition is different. Biomass ash usually has a higher melting point. Depending on the melting temperature of the ash, a slagging agent – which lowers the melting point of the ash – must be added to ensure a good melt of the ash.

Table 11
Sweet syngas characteristics.

	Unit	EP	TOPS	Coal
H_2O	mol% wet	1	1	0
H_2	mol% dry	68	68	68
CO	mol% dry	29	30	30
CO_2	mol% dry	0	0	0
Other	mol% dry	2	2	2
Energy density	MJ_{HHV}/Nm^3	12	12	12
Gas volume flow	Nm^3/s	42	51	64

Biomass – not TOPS – is much more fibrous than coal, requiring a hammer mill instead of crusher. However, biomass can use the more efficient hydraulic piston with screw feeding system. It is therefore appropriate to use a dedicated pre-treatment train for EP and a different train for coal, both with their own pressurised storage bins. TOPS can be used in either feeding system, although the biomass feeding system is more efficient (Table 3).

The higher the biomass fraction, the higher the moisture content in the raw syngas. This reduces water consumption of the wet scrubber as the cooling – and resulting condensation of the moisture – already results in the water flows needed to remove the halides.

As known, coal is sulphur-rich, while TOPS and EP are sulphur-poor. This results in a reduction in sulphur content in the syngas when switching from coal to biomass. Also, the lower CGE results in a higher CO_2 content: coal derived syngas entering the AGR contains 5300 ppm H_2S and 21% CO_2; TOPS derived syngas contains <100 ppm H_2S and 23% CO_2; EP derived syngas contains <100 ppm H_2S and 25% CO_2. The low sulphur flows at high biomass fractions causes problems in the Claus unit. Therefore, sulphur needs to be recycled.

By altering the recirculation rates of the Rectisol, the sweet syngas leaving the AGR is almost similar in composition, regardless of the feedstock (Table 11). Total sweet syngas flow does differ significantly. Therefore, downstream conversion sections only have to cope with a change in gas flow and not with a change in gas composition.

Modern gas turbines are tuned for a specific gas composition and flow. Flexibility means that they must handle different syngas compositions and flows. Of special interest are the syngas H_2:CO ratio, energy density and flow. The H_2:CO ratio varies significant when altering methanol production. For FT-liquids and urea production, this variation is limited (Table 12).

6.1. Reference cases

The overall energy efficiency and carbon balances of the extreme cases – pure biomass or pure coal – without substitution are given in Table 13. Using biomass results in slightly lower efficiencies compared to using coal. When using TOPS this difference is even smaller than when using EP. Overall efficiencies of producing methanol are relatively low. The main reason is the mediocre conversion of the liquid phase methanol synthesis process.[4] This, combined with a

[3] A possible alternative is to over-dimension the gasifier to maintain a constant chemical energy input. In that case the gasifier is operated at part-load when using coal and at full load when using biomass.

[4] The LPMeOH process was selected as the improved heat management allows the use of CO-rich syngas. It also improves the conversion compared to the conventional

Table 12
Gas turbine characteristics when changing feedstock from pure biomass to pure coal.

		Electricity	FT-liquids	Methanol	Urea
H_2:CO ratio syngas to GT					
Reference	Coal, TOPS and EP	2.3	4.3	12.6	2.8
Constant chemical production	TOPS → coal	N.A.	2.9 → 4.3	7.6 → 12.7	2.6 → 2.8
	EP → coal	N.A.	2.6 → 4.2	5.9 → 12.6	2.5 → 2.8
Energy density syngas to GT before N_2 dilution (MJ_{LHV}/Nm^3)					
Reference	Coal, TOPS and EP	11	11	9	8
Constant chemical production	TOPS → coal and EP → coal	N.A.	11	9	8
Energy flow syngas to GT (MW_{LHV})					
Reference	Coal	1428	259	690	721
	TOPS	1148	203	550	580
	EP	908	161	436	459
Constant chemical production	TOPS → coal	N.A.	203 → 483	550 → 824	580 → 854
	EP → coal	N.A.	161 → 681	436 → 944	459 → 979

Table 13
Overall energy efficiencies of XtY facilities.

Energy efficiency		Coal	TOPS[a]	EP
Electricity (XtP)	Power	40%	39% (35%)	38%
FT-liquids (XtL)	FT	49%	47% (42%)	43%
	Power	10%	11% (10%)	12%
	Total	60%	58% (53%)	55%
Methanol (XtM)	MeOH	33%	31% (28%)	29%
	Power	21%	21% (19%)	21%
	Total	53%	52% (47%)	49%
Urea (XtU)	Urea	29%	28% (25%)	25%
	Power	22%	22% (20%)	22%
	Total	51%	50% (45%)	47%

[a] During torrefaction roughly 10% of the biomass energy is lost. The values in bracket include this penalty.

once-through concept, results in a relatively low methanol and high electricity production. The low conversion efficiency of the electricity production results in a low overall efficiency for the XtM facilities.

The allocation of carbon is displayed in Fig. 8. As the syngas is shifted to a specific H_2:CO ratio, regardless of production, the amount of CO_2 captured is identical for each feedstock. This also limits the capture rate. The IG-PG facility is not optimised for CO_2 capture, but for flexibility. As a result, between 56% and 65% of the carbon is captured. Only with urea production, a part of the captured CO_2 is used and gets embedded in the product. For the FT-liquids and methanol production the carbon in the syngas is used and not the captured CO_2. Therefore, with urea production less CO_2 needs to be sequestered. The fraction of carbon that is captured increases from coal < TOPS < EP. The reason for this is that a higher fraction of the feedstock is combusted in the gasifier when using biomass as feedstock. CO_2 emissions vary depending on feedstock used and commodity produced. This is because a part of the carbon ends up in the chemical end-products for the FT-liquids and methanol cases. Therefore, the XtP and XtU cases have the largest emissions. As the syngas to FT-liquids conversion is higher than syngas to methanol, the XtL cases have the lowest emissions.

6.1.1. Electricity

In the following section the results of electricity production with and without flexibility are given. Also, the effects of feedstock substitution are presented.

6.1.1.1. Reference. Total output drops from 800 MW_e when using pure coal to 668 MW_e (−16%) when using pure TOPS and 550 MW_e (−31%) when using pure EP (Fig. 9). This is caused mainly by the lower thermal input when using biomass: thermal input is 2000 MW_{th} for pure coal, but 1696 MW_{th} for pure TOPS and 1464 MW_{th} for pure EP. Total efficiency is hardly affected by the different feedstocks. It drops from 40% for pure coal to 39% for pure TOPS and 38% for pure EP. The net CO_2 emissions of the IG-CC with CO_2 capture drop from 343 kg CO_2/MW h for pure coal to −506 kg CO_2/MW h for pure TOPS and −589 kg CO_2/MW h for pure EP. CO_2-neutral electricity is produced at biomass energy fractions of 41% (48 wt%) for TOPS and 39% (50 wt%) for EP. Although the gasifier has a larger feedstock mass input when using biomass compared to coal, total carbon input is lower for the biomass cases as the carbon fraction in coal is much higher. The result is a slight reduction in absolute amount of CO_2 captured. The summarised mass and energy balances are given in Tables 14–16.

6.1.1.2. Feedstock substitution. When the facility is sized for pure coal, but biomass is used, the gas turbine operates at part-load. However, the inlet temperature can be maintained at the desired level of 1375 °C. Although the efficiency of the gas turbine is reduced, total electricity output remains almost constant (Fig. 9). The largest difference occurs at the substitution of coal for EP. The reduction in electric output compared to the EP-IGCC base case is 13 MW_e or 2%. There are three main reasons why this reduction is so small. First, a less efficient gas turbine results in a hotter flue gas. This improves the electricity production from the steam cycle. Second, the GT can reduce the air flow, thereby keeping the flue gas at a high temperature. Third, the gas turbine produces a smaller fraction of the total output than a standard NGCC. This is due to the additional heat streams inherent to an IGCC.

6.1.2. FT-liquids

In the following section the results of FT-liquids production with and without flexibility are given. Also, the effects of feedstock and production substitution are presented.

6.1.2.1. Reference. Overall plant output (FT-liquids and electricity) drops from 1198 MW using pure coal to 990 MW with pure TOPS and 804 MW with pure EP (Fig. 10). Besides this drop in output, also the ratio FT-liquids:electricity changes; a low ratio means a relatively low chemical and high electricity output. Increasing the biomass fraction reduces this ratio. This effect is larger for EP compared to TOPS. Total efficiency drops from 60% for pure coal to 58% for pure TOPS and 55% for pure EP. Net CO_2 emissions drop from 44 t CO_2/h with pure coal to −523 t CO_2/h with pure TOPS and −469 t CO_2/h with pure EP. CO_2 neutral production is achieved at co-feeding rates of 7% TOPS or 6% EP on an energy basis. This does not include the carbon bound in the FT-liquids. If the carbon in the FT-liquids counts as GHG emission, CO_2-neutral FT-liquids production needs the same biomass fraction as electricity production

process, although conversion remains mediocre. Finally, the LPMeOH process has been tested with variable production.

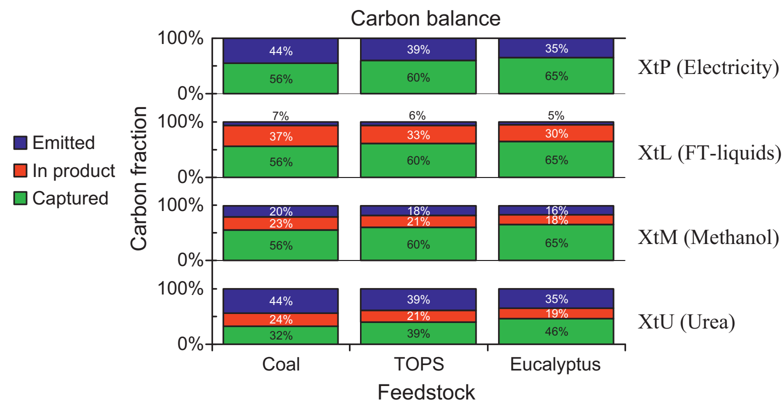


Fig. 8. Carbon distribution. The numbers display the carbon fractions.

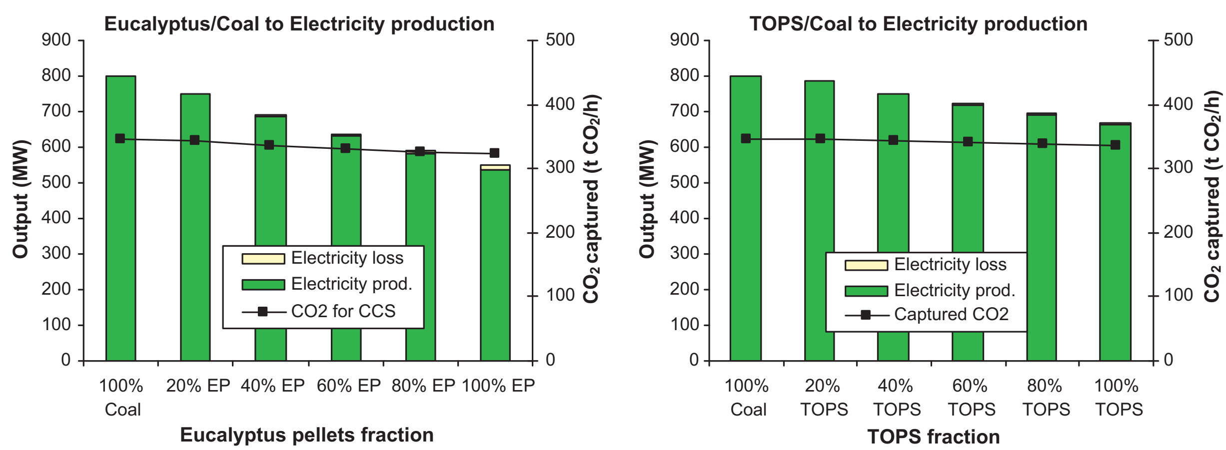


Fig. 9. Effect different feedstocks on power production.

Table 14
Summary mass balances XtP reference cases.

Biomass fraction	Unit	TOPS			Coal	EP		
	%	100	60	20	0	20	60	100
Biomass	t/h	283	182	65		71	186	274
Coal	t/h		86	185	236	178	78	
Oxygen gasifier	t/h	168	183	200	208	199	181	168
Syngas to power	t/h	173	190	211	216	201	166	139
Air to GT	t/h	3.37k	3.71k	4.1k	4.19k	3.89k	3.2k	2.67k
Flue gas	t/h	3.77k	4.14k	4.58k	4.69k	4.34k	3.58k	2.98k
CO_2 emitted	t CO_2/h	220	242	268	274	254	209	174

(41% for TOPS and 39% for EP). The summarised mass and energy balances are given in Tables 17–19.

6.1.2.2. Feedstock substitution. If coal is substituted by biomass, the ratio FT-liquids:electricity is kept constant and part-load effects are included, the worst case – substituting pure coal for pure EP – has a 3 MW_e drop in electric output compared to the pure EP reference BtL case.

The effect of constant FT-liquid production and substituting biomass for coal is displayed in Fig. 11 and Table 20. The TOPS cases are displayed on the left and the EP cases on the right. FT-liquid production is set at either maximum FT-liquid production at pure

Table 15
Summary energy balances XtP reference cases.

Biomass fraction	Unit	TOPS			Coal	EP		
	%	100	60	20	0	20	60	100
Biomass input	MW_{HHV}	1696	1091	392		378	990	1464
Coal input	MW_{HHV}		727	1569	2000	1513	660	
Total internal power use	MW_e	−146	−156	−168	−172	−164	−148	−135
Net power production	MW_e	668	722	785	800	749	637	550
Total efficiency	$\%_{HHV}$	39.4	39.7	40.0	40.0	39.6	38.6	37.6

Table 16
Summary carbon balances XtP reference cases.

Biomass fraction	Unit	TOPS			Coal	EP		
	%	100	60	20	0	20	60	100
Carbon input from biomass	t C/h	152	98	35		35	92	136
Carbon input from coal	t C/h		62	133	169	128	56	
Carbon captured	t C/h	92	93	95	94	94	90	88
Carbon emitted	t C/h	60	66	73	75	69	57	48
Fraction C captured	%	60	58	56	56	57	61	65
Fraction C in flue gas	%	39	41	43	44	42	39	35
Net CO_2 emission	t C/h	−92	−32	38	75	34	−35	−88
CO_2 emissions	kg CO_2/MWh	−504	162	177	343	167	−200	−589
CO_2 emissions	kg CO_2/GJ$_e$	−141	−45	49	95	47	−56	−164

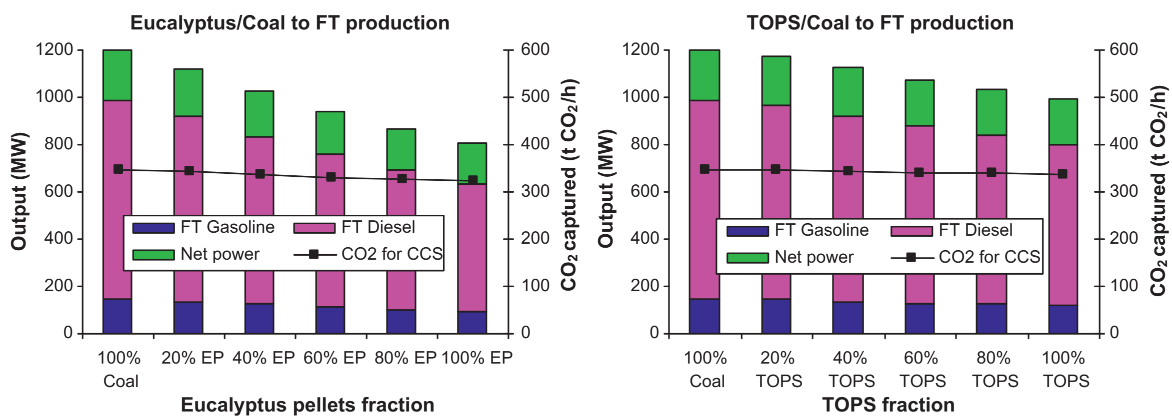


Fig. 10. Effect different feedstocks on FT-liquids production.

Table 17
Summary mass balances XtL reference cases.

Biomass fraction	Unit	TOPS			Coal	EP		
	%	100	0.6	0.2	0	0.2	0.6	1.0
Biomass	t/h	283	182	65		71	186	274
Coal	t/h		86	185	236	178	78	
Oxygen gasifier	t/h	168	183	200	208	199	181	168
Syngas to power	t/h	3	3	3	3	3	2	2
Air to GT	t/h	0.61k	0.67k	0.75k	0.77k	0.71k	0.58k	0.48k
Flue gas	t/h	0.68k	0.75k	0.83k	0.87k	0.79k	0.65k	0.54k
FT-liquids	t/h	61	67	74	75	70	58	48
CO_2 emitted	t CO_2/h	31	34	38	40	36	30	25

biomass or at a 50/50 coal:biomass ratio on an energy basis. The reference CtL case is displayed in the middle of both graphs. As can been seen, power production is increased when coal is used. Also, CO_2 emissions are higher. Compared to the reference cases, CO_2 capture rates are not affected by the substitution. This is because the AGR is adjusted to capture a specific fraction (Table 9). As FT-liquid production is more efficient than electricity production, efficiency drops when using coal. The modelled gas turbine has good part-load behaviour down to 40%. Only the pure EP case demands a lower load.

6.1.2.3. Production substitution. With pure coal as feedstock and switching from FT-liquids production to electricity production reduces overall conversion efficiency from 59% to 40%. The difference between electricity output in full FT-production and the CtL base case is 18 MW$_e$. For this extreme case, the gas turbine

Table 18
Summary energy balances XtL reference cases.

Biomass fraction	Unit	TOPS			Coal	EP		
	%	100	60	20	0	20	60	100
Biomass input	MW$_{HHV}$	1696	1091	392		378	990	1464
Coal input	MW$_{HHV}$		727	1569	2000	1513	660	
Total internal power use	MW$_e$	−143	−152	−163	−167	−160	−144	−132
Net power production	MW$_e$	192	198	206	210	200	185	173
FT-gasoline production	MW$_{HHV}$	118	130	144	147	137	113	94
FT-diesel production	MW$_{HHV}$	681	748	825	841	784	645	537
Total efficiency	%$_{HHV}$	58.4	59.2	59.9	59.9	59.3	57.1	54.9

Table 19
Summary carbon balances XtL reference cases.

Biomass fraction	Unit	TOPS			Coal	EP		
	%	100	60	20	0	20	60	100
Carbon input from biomass	t C/h	152	98	35		35	92	136
Carbon input from coal	t C/h		62	133	169	128	56	
Carbon in FT-liquids	t C/h	51	56	62	63	59	48	40
Carbon captured	t C/h	92	93	95	94	94	90	88
Carbon emitted	t C/h	9	10	11	11	10	8	7
Fraction C captured	%	60	58	56	56	57	61	65
Fraction C in products	%	33	35	37	37	36	33	30
Fraction C in flue gas	%	6	6	6	7	6	6	5
Net CO_2 emission	t C/h	−143	−88	−24	12	−24	−83	−128
CO_2 emissions chemical[a]	kg CO_2/GJ	−206	−124	−46	−8	−48	−135	−234

[a] Exported electricity is credited for 95 kg CO_2/GJ_e. This is subtracted from the carbon emissions.

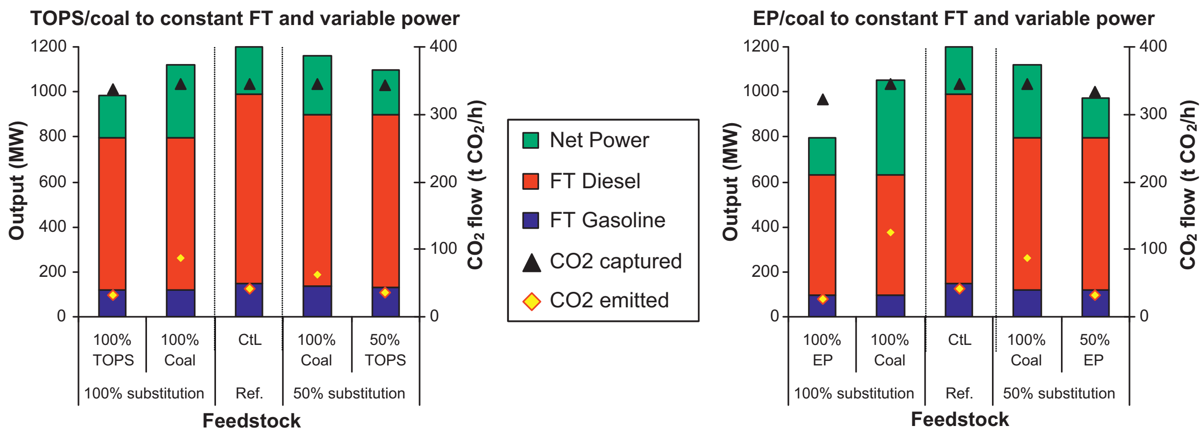


Fig. 11. Impact of feedstock substitution with constant FT-liquids production.

Table 20
Effect coal:biomass substitution with constant FT-liquid production.[a]

	Substitution degree	FT-prod. (MW_{th})	Power prod. (MW_e)	CO_2 emissions (t CO_2/h)	Total efficiency	GT load
TOPS	0% → 50%	899	263 → 197	63 → 37	58% → 59%	64%
	0% → 100%	799	323 → 184	87 → 32	56% → 58%	43%
EP	0% → 50%	794	326 → 181	88 → 32	56% → 57%	42%
	0% → 100%	631	423 → 163	126 → 23	53% → 54%	24%

[a] Before the arrow represents feeding only coal. After the arrow represents (partly) substitution of coal by biomass.

operates at 19% part-load, which is very inefficient. The amount of CO_2 captured remains constant but CO_2 emissions increase as less CO_2 is bound in the FT-liquids. This increase is from 42 t CO_2/h to 274 t CO_2/h. The relation between FT-liquids:electricity ratio and output is displayed in Fig. 12. Similar results have been found for TOPS and EP.

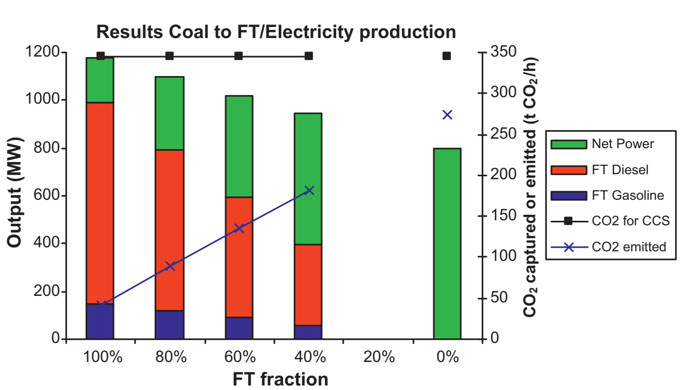


Fig. 12. Impact substituting FT-production for power production.

6.1.3. *Methanol*

In the following section the results of methanol production with and without flexibility are given. Also, the effects of feedstock and production substitution are presented.

6.1.3.1. Reference. Overall plant output drops from 1066 MW using pure coal to 886 MW with pure TOPS and 724 MW with pure EP (Fig. 13). The ratio methanol:electricity decreases with increasing biomass fraction. The effect is larger for EP than for TOPS. Total efficiency drops from 53% for pure coal to 52% for pure TOPS and 49% for pure EP. Net CO_2 emissions drop from 133 t CO_2/h with pure coal to −452 t CO_2/h with pure TOPS and −414 t CO_2/h with pure EP. This does not include the carbon bound in the methanol. If the carbon in the methanol counts as GHG emission, CO_2-neutral methanol production needs the same biomass fraction as electricity production (41% for TOPS and 39% for EP). If the carbon in the methanol is not considered as GHG emission, CO_2 neutral production is achieved at co-feeding rates of 20% TOPS or 19% EP on an energy basis. The summarised mass and energy balances are given in Tables 21–23.

6.1.3.2. Feedstock substitution. If coal is substituted by biomass, the ratio methanol:electricity is kept constant and part-load effects are included, the worst case – substituting pure coal for pure EP – has

Table 21
Summary mass balances XtM reference cases.

Biomass fraction	Unit	TOPS			Coal	EP		
	%	100	60	20	0	20	60	100
Biomass	t/h	283	182	65		71	186	274
Coal	t/h		86	185	236	178	78	
Oxygen gasifier	t/h	168	183	200	208	199	181	168
Syngas to power	t/h	83	91	101	104	96	79	66
Air to GT	t/h	1.6k	1.77k	1.95k	2.01k	1.85k	1.53k	1.27k
Flue gas	t/h	1.77k	1.95k	2.16k	2.22k	2.04k	1.68k	1.4k
Methanol	t/h	84	92	102	103	96	80	67
CO_2 emitted	t CO_2/h	99	108	120	124	114	94	78

Table 22
Summary energy balances XtM reference cases.

Biomass fraction	Unit	TOPS			Coal	EP		
	%	100	60	20	0	20	60	100
Biomass input	MW_{HHV}	1696	1091	392		378	990	1464
Coal input	MW_{HHV}		727	1569	2000	1513	660	
Total internal power use	MW_e	−140	−149	−160	−163	−156	−141	−130
Net power production	MW_e	357	380	407	415	391	342	304
MeOH production	MW_{HHV}	529	580	639	651	607	502	420
Total efficiency	$\%_{HHV}$	52.3	52.8	53.3	53.3	52.8	51.1	49.5

Table 23
Summary carbon balances XtM reference cases.

Biomass fraction	Unit	TOPS			Coal	EP		
	%	100	60	20	0	20	60	100
Carbon input from biomass	t C/h	152	98	35		35	92	136
Carbon input from coal	t C/h		62	133	169	128	56	
Carbon in methanol	t C/h	32	35	38	39	36	30	25
Carbon captured	t C/h	92	93	95	94	94	90	88
Carbon emitted	t C/h	27	30	33	34	32	26	22
Fraction C captured	%	60	58	56	56	57	61	65
Fraction C in products	%	21	22	23	23	22	20	18
Fraction C in flue gas	%	18	19	20	20	19	18	16
Net CO_2 emission	t C/h	−123	−66	−0	36	−2	−64	−113
CO_2 emissions chemical[a]	kg CO_2/GJ	−305	−182	−64	−7	−67	−198	−345

[a] Exported electricity is credited for 95 kg CO_2/GJ_e. This is subtracted from the carbon emissions.

a 6 MW_e drop in electric output compared to the pure EP reference BtM case.

The effect of constant methanol production and substituting biomass for coal is displayed in Fig. 14 and Table 24. Similar effects as those occurring with the FT-liquids cases can be seen, although less severe. The smaller impact is a result of the poor conversion of the LPMeOH synthesis.[5] This, combined with a once-through concept, results in large amounts of unreacted syngas directed to the gas turbine. This results in an already significant load for the gas turbine. Compared to the FT-liquids cases, the additional syngas to the gas turbine – caused by the substitution of biomass for coal – absolutely similar, but relatively much smaller.

6.1.3.3. Production substitution. Substituting production from methanol to electricity gives similar results as with the substitution of FT-liquids and electricity. Overall efficiency drops from 52% for methanol production to 40% for only electricity production. The difference between electricity output in full methanol production and the CtM base case is 20 MW_e. For this extreme case, the gas turbine operates at 48% part-load. CO_2 emissions increase from 126 t CO_2/h to 274 t CO_2/h. The relation between methanol:electricity ratio and output is displayed in Fig. 15.

6.1.4. Urea

In the following section the results of urea production with and without flexibility are given. Also, the effects of feedstock and production substitution are presented.

6.1.4.1. Reference. Overall plant output drops from 1015 MW using pure coal to 841 MW with pure TOPS and 686 MW with pure EP (Fig. 16). The ratio urea:electricity decreases with increasing biomass fraction. The effect is larger for EP than for TOPS. Total efficiency drops from 51% for pure coal to 50% for pure TOPS and 47% for pure EP. Net CO_2 emissions drop from 421 t CO_2/h with pure coal to −220 t CO_2/h with pure TOPS and −229 t CO_2/h with pure EP. This does not include the carbon bound in the urea. If the carbon in the urea counts as GHG emission, CO_2-neutral urea production needs the same biomass fraction as electricity production (41% for TOPS and 39% for EP). If the carbon in the urea is not considered as GHG emission, CO_2 neutral production is achieved at co-feeding rates of 62% TOPS or 58% EP on an energy basis. The summarised mass and energy balances are given in Tables 25–27.

6.1.4.2. Feedstock substitution. If coal is substituted by biomass, the ratio urea:electricity is kept constant and part-load effects are

[5] This process was selected for its superior heat management and part-load performance and its lower capital costs compared to the conventional methanol synthesis process.

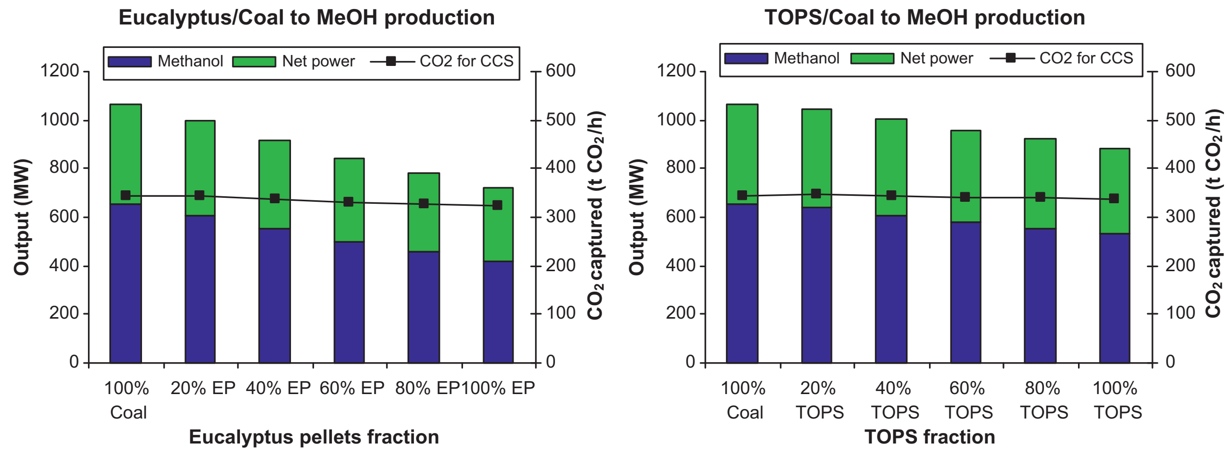


Fig. 13. Effect different feedstocks on methanol production.

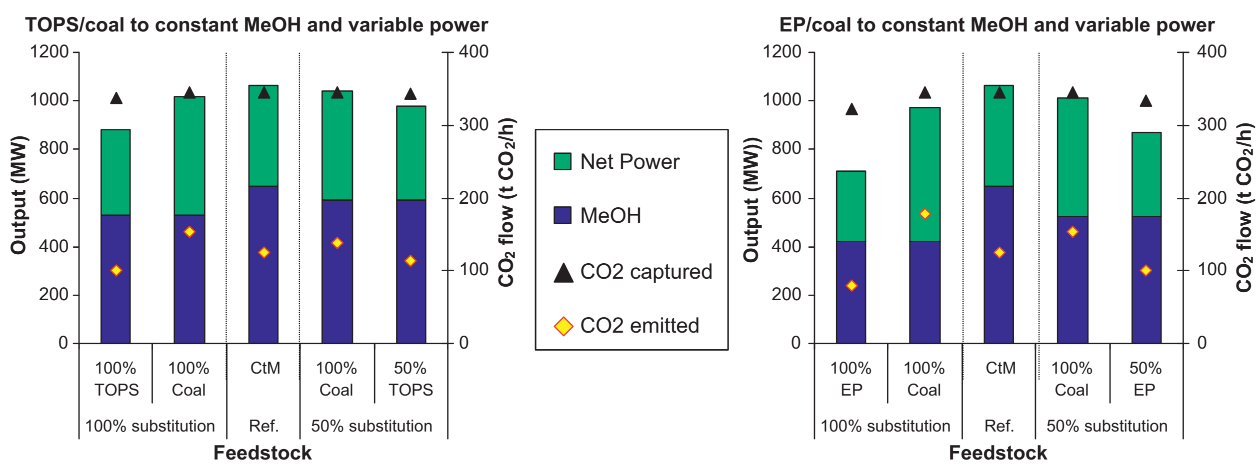


Fig. 14. Impact of feedstock substitution with constant methanol production.

Table 24
Effect coal:biomass substitution with constant methanol production.[a]

	Substitution degree	MeOH prod. (MW_{th})	Power prod. (MW_e)	CO_2 emissions ($t\,CO_2/h$)	Total efficiency	GT load
TOPS	0% → 50%	594	449 → 384	139 → 113	52% → 53%	82%
	0% → 100%	529	487 → 350	154 → 100	51% → 52%	66%
EP	0% → 50%	526	489 → 346	154 → 100	51% → 51%	66%
	0% → 100%	420	552 → 290	178 → 79	49% → 49%	46%

[a] Before the arrow represents feeding only coal; after the arrow represents (partly) substitution of coal by biomass.

included, the worst case – substituting pure coal for pure EP – has a 7 MW_e drop in electric output compared to the pure EP reference BtU case.

The effect of constant urea production and substituting biomass for coal is displayed in Fig. 17 and Table 28. Urea production does not differ that significantly. The reason for this is that in the base cases significant amounts of syngas are already directed to the gas turbine to maintain a sufficient energy density in the gas turbine. In all cases the gas turbine operates above the 40% load limit.

Table 25
Summary mass balances XtU reference cases.

Biomass fraction	Unit	TOPS			Coal	EP		
	%	100	60	20	0	20	60	100
Biomass	t/h	283	182	65		71	186	274
Coal	t/h		86	185	236	178	78	
Oxygen gasifier	t/h	168	183	200	208	199	181	168
Syngas to power	t/h	220	242	269	275	255	211	176
Air to GT	t/h	1.72k	1.89k	2.09k	2.14k	1.98k	1.63k	1.36k
Flue gas	t/h	1.94k	2.14k	2.36k	2.42k	2.24k	1.85k	1.54k
Urea	t/h	160	176	195	200	185	152	127
CO_2 emitted	$t\,CO_2/h$	218	240	265	271	251	207	172

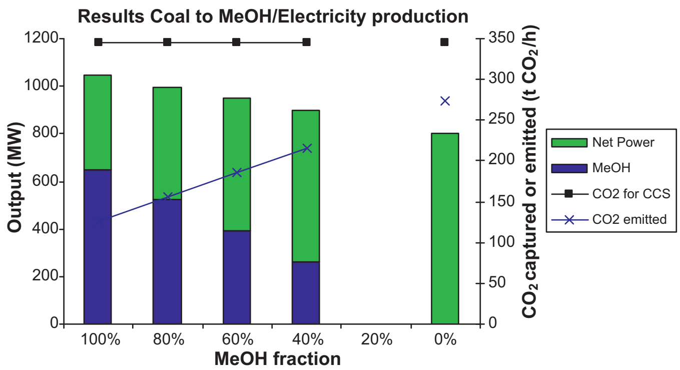


Fig. 15. Impact substituting methanol production for power production.

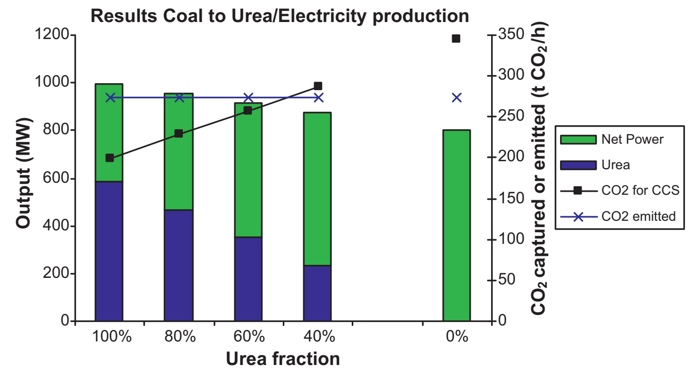


Fig. 18. Impact substituting urea production for power production.

6.1.4.3. *Production substitution.* When substituting urea production for electricity production overall efficiency drops from 50% for urea production to 40% for only electricity production. The difference between electricity output in full urea production and the CtU base case is 19 MW_e. For this extreme case, the gas turbine operates at 52% part-load. CO_2 emissions remain constant at 274 t CO_2/h. But, as less captured CO_2 is used for urea production, the amount of CO_2 that needs to be stored increases from 199 t CO_2/h to 345 t CO_2/h. The relation between urea:electricity ratio and output is displayed in Fig. 18.

7. Discussion

7.1. Model validation

A drawback of our modelling studies is that results have not been tested. To validate the model two different parameters were compared with public available literature sources. The first is parameter is the CGE of the gasifier. The obtained CGE for coal (81%) matches the value reported in literature (79–83%) [16,53,54]. For EP and TOPS no CGE values using EF gasifiers were found in open literature.

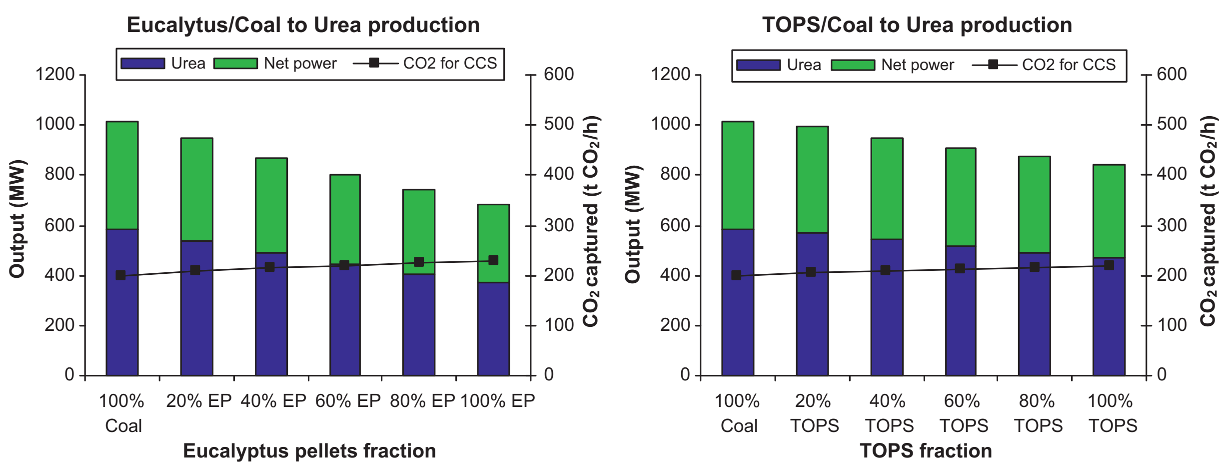


Fig. 16. Effect different feedstocks on urea production.

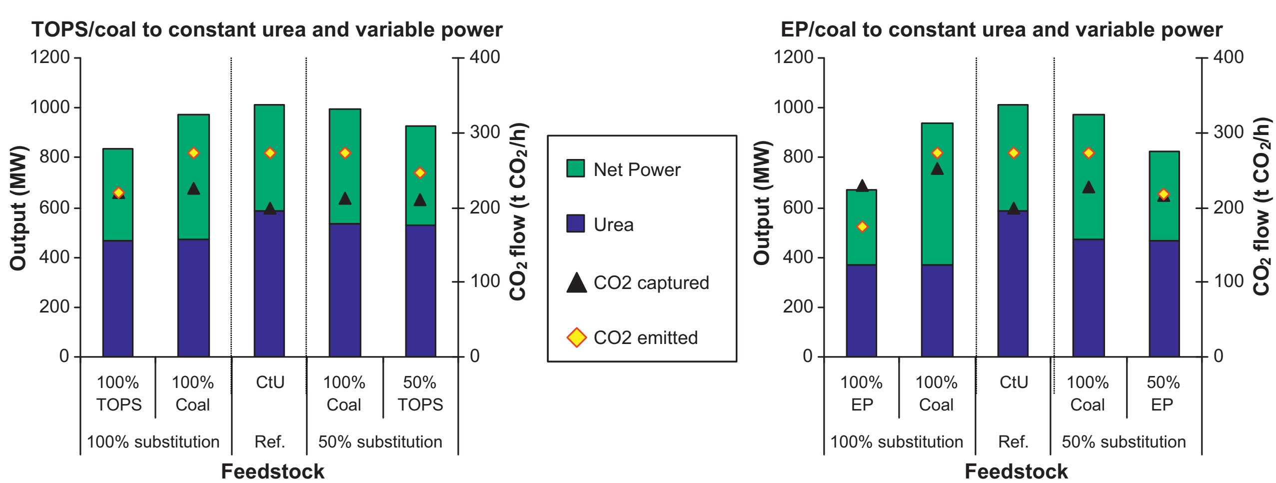


Fig. 17. Impact of feedstock substitution with constant urea production.

Table 26
Summary energy balances XtU reference cases.

	Unit	TOPS			Coal	EP		
Biomass fraction	%	100	60	20	0	20	60	100
Biomass input	MW_{HHV}	1696	1091	392		378	990	1464
Coal input	MW_{HHV}		727	1569	2000	1513	660	
Total internal power use	MW_e	−156	−167	−180	−184	−176	−157	−143
Net power production	MW_e	372	397	425	431	408	356	316
Net power production	MW_e	468	515	569	583	540	444	370
Total efficiency	$\%_{HHV}$	49.6	50.1	50.7	50.7	50.1	48.5	46.9

Table 27
Summary carbon balances XtU reference cases.

Biomass fraction	Unit	TOPS			Coal	EP		
	%	100	60	20	0	20	60	100
Carbon input from biomass	t C/h	152	98	35		35	92	136
Carbon input from coal	t C/h		62	133	169	128	56	
Carbon in urea	t C/h	32	35	39	40	37	30	25
Carbon captured	t C/h	60	58	56	54	57	60	63
Carbon emitted	t C/h	60	66	73	75	69	57	47
Fraction C captured	%	39	36	33	32	35	41	46
Fraction C in products	%	21	22	23	24	23	21	19
Fraction C in flue gas	%	39	41	43	44	42	38	35
Net CO_2 emission	t C/h	−92	−32	38	75	34	−35	−88
CO_2 emissions chemical[a]	kg CO_2/GJ	−277	−137	−4	60	−8	−157	−324

[a] Exported electricity is credited for 95 kg CO_2/GJ_e. This is subtracted from the carbon emissions.

The second parameter is overall plant efficiencies. For coal and EP these are slightly higher than open literature values (see Table 29). Values for TOPS using EF gasifiers were not found. Possible explanations are that the modelled IG-PG facility only captures part of the CO_2 – thereby reducing the CO_2 capture penalty – and the use of a different gas turbine than commonly used in the literature sources. Sometimes a different coal or biomass type is used. This can also result in different overall efficiencies. A possible explanation of the difference between the XtU efficiencies is that the literature value is based on a overall conversion efficiency, while the model value is based on the performance of the different individual components.

7.2. Fuel flexibility

Switching from coal to TOPS or EP has some important impacts on feedstock volumes, oxygen consumption and syngas composition and volume:

- The difference in hardness and fibrousness of coal and biomass can cause feeding problems when using one feeding system. To prevent these problems and increase flexibility separate feeding systems and pressurised bins for feedstock storage should be used. This also allows the use of more efficient grinding and pressurising techniques.
- At this time no EF gasifier exists that allows co-firing biomass with a higher fraction than 50% on a mass basis, equivalent to 30% on an energy basis. Main bottlenecks are the low heating value of the biomass and the fouling (see comment below). It is questionable whether the current generation EF gasifiers can co-fire high biomass fractions, but, especially with TOPS, 50% co-firing on an energy basis is technically possible, although it might be uneconomical.
- Oxygen consumption drops by 10%. This holds both for substituting coal for EP or TOPS. Current generation ASU have no problems with this load reduction.
- The differences in coal slag and biomass slag behaviour can create problems. Especially the reduction in slag amount and the increase in slag melting temperature are problematic. Too little slag can result in a loss of protective slag layer in the gasifier. Adding flux material and recycling slag, by crushing and adding slag can prevent this. It does mean constant monitoring of the slag layer.
- The differences in syngas composition (Table 10) can be adjusted and levelled. Therefore, the chemical conversion sections only have to cope with a difference in syngas flow and not in syngas composition.
- The Claus unit will give problems as pure biomass will reduce the sulphur flow to 1% compared to coal. This is much lower than the minimal requirement of the Claus unit. To prevent operational problems, sulphur may need to be recycled. This will slightly lower the total energy efficiency for wood and TOPS.
- Not modelled is the fouling in the syngas cooler. Although the fouling is negligible compared to the main mass and energy flows, the cumulative effect of prolonged fouling is important as this decreases the efficiency and availability of the syngas cooler. The higher chloride concentration in biomass will increase fouling compared to coal [52]. To limit this problem biomass fractions are currently restricted to maximum 50 wt% and the quench temperature is lowered from 900 °C to 850 °C.

Table 28
Effect coal:biomass substitution with constant urea production.[a]

	Substitution degree	Urea prod. (MW_{th})	Power prod. (MW_e)	CO_2 emissions (t CO_2/h)	Total efficiency	GT load
TOPS	0% → 50%	532	463 → 401	273 → 247	50% → 50%	84%
	0% → 100%	468	500 → 366	274 → 219	49% → 49%	68%
EP	0% → 50%	471	502 → 361	274 → 218	49% → 49%	68%
	0% → 100%	370	566 → 302	274 → 174	47% → 46%	47%

[a] Before the arrow represents feeding only coal; after the arrow represents (partly) substitution of coal by biomass.

Table 29
Overall system efficiency comparison between model and literature.

	Literature			Aspen model		
	Coal	TOPS	EP	Coal	TOPS	EP
XtP	32–40[a]	N.A.	36 [5]	40	39	38
XtL	57 [85]	N.A.	53 [5]	60	58	55
XtM	51 [86] (LHV)	N.A.	N.A.	53	52	49
XtU	44 [19]	N.A.	N.A.	51	50	47

[a] The first value includes CO_2 capture. The second value excludes CO_2 capture.

- Changing the gas turbine fuel composition will result in different NO_x emissions. These emissions have not been modelled, but must be kept low enough to fulfil environmental legislation.

7.3. Model assumptions

Several assumptions were made to model flexible IG-PG facilities (5). This led to the configuration displayed in Fig. 6, but other configurations are also possible:

- In this study the limiting factor for the scale of the facility was the gasifier. It dictated the feedstock input and thus mass and energy balances throughout the facility. Commonly, a facility with gas turbines makes the gas turbines the limiting factor – the facility is dimensioned on the gas turbine. This ensures that the gas turbines are always operated at optimal efficiency. By making the gasifier the limiting factor, the gasifier and gas cleaning train, which are the most capital-intensive part of an IG-PG facility, were optimally used. Also, by having a constant gas stream exiting the gasifier, the effects of flexibility are much clearer and more transparent.
- Alternatively, if the gasifier is the limiting factor, natural gas can be co-fed in the gas turbine to keep the gas turbine at optimal efficiency.
- The H_2:CO ratio was kept constant at 2.3 in the sweet syngas. This was done to keep the gas cleaning and optimising operating as constant as possible. Normally, CCS facilities will try to capture as much CO_2 as possible. Allowing the H_2:CO ratio in the sweet syngas to change can reduce CO_2 emissions to almost zero.
- The model used a sour and a sweet shift, with the AGR in between. Alternatively the train could be: S-capture, sweet shift, CO_2 capture.

8. Conclusion

In this study the technical possibilities and impacts on overall efficiency and carbon balance of introducing flexibility to IG-PG facilities was investigated. An IG-PG facility consists of numerous specific components. Of each component the technical performance at full and part-load was investigated. This data was combined with different IG-PG facility configurations in an AspenTech process flowsheet. This allowed the calculation of the different case studies by altering the used feedstock or configuration. By analysing the resulting mass and energy balances the effects of flexibility on IG-PG facilities were determined.

Electricity with CCS can be produced at $40\%_{HHV}$ efficiency. Using TOPS or EP lowers this to $39\%_{HHV}$ and $38\%_{HHV}$ respectively. Of the carbon in the feedstock 56–65%, for coal and EP respectively, is captured by CO_2 removal. This value is much lower than the normally reported 90% capture for IGCC-CCS facilities, due to a much higher CO-concentration in the sweet syngas as the facility is designed for flexibility. The CO is not captured by the acid gas removal. It is combusted in the gas turbine and the carbon is emitted with the flue gas as CO_2.

When using pure EP in a full feedstock flexible IG-PG facility lowers total output by 13 MW_e or 2% of overall efficiency compared to the output when using a static IG-PG facility due to part-load penalties.

FT-liquids (and electricity) with CCS can be produced at $55–60\%_{HHV}$ efficiency, with the lower value corresponding to pure EP as feedstock and the higher value to pure coal. Again, between 56% and 65% of the carbon in the feedstock is captured by CO_2 removal. But, an additional 37–30% of the carbon ends up in the FT-liquid. The carbon fraction emitted with the flue gas is 5–7% for pure coal and pure EP respectively.

When using pure EP in a full feedstock flexible IG-PG facility lowers total output by 3 MW_e or < 1% of overall efficiency compared to the output of a static IG-PG facility due to part-load penalties.

Variable feedstock with constant chemical production results in a too small gas turbine load for the case where pure EP is substituted by pure coal. The load is reduced from 100% with coal to only 24% with EP. The smallest load reduction is achieved with 50/50 TOPS. Here the gas turbine is reduced to 64% load.

Complete substitution of FT-liquids production for electricity production is not straight forward as gas turbine load is reduced to 18%. Possible solutions are adding natural gas as make-up fuel or having multiple smaller gas turbines. Even when FT-liquids production is varied between 60% and 100%, the gas turbine load still decreases to 36%.

Methanol production (and electricity) with CCS has similar results as FT-liquids, although less extreme. Overall efficiency is between $49\%_{HHV}$ and $53\%_{HHV}$. Total carbon emissions are between 16% and 20%.

When using pure EP in a full feedstock flexible IG-PG facility lowers total output by 6 MW_e or <1% of overall efficiency compared to the output of a static IG-PG facility due to part-load penalties.

Variable feedstock with constant chemical production does not exceed part-load limits. In the XtM cases the gas turbine already has a higher load compared to the XtL cases. Therefore, although the absolute difference in gas turbine duty is similar to the XtL cases, the gas turbine load is reduced to only 46% when substituting coal for biomass.

Complete substitution of methanol production for electricity production is possible. Gas turbine load is reduced to 48%.

Urea production (and electricity) with CCS has overall efficiencies of $47–51\%_{HHV}$. Total carbon emissions are between 35% and 44%. For urea production only carbon in the captured CO_2 is used. The remaining carbon – mostly in the form of CO – is directed to the gas turbine, combusted and emitted.

When using pure EP in a full feedstock flexible IG-PG facility lowers total output by 7 MW_e or <1% of overall efficiency compared to the output of a static IG-PG facility due to part-load penalties.

Similar to the methanol cases, variable feedstock with constant chemical production is possible. In the most extreme case gas turbine load is reduced to 47%.

The normal XtU cases have an even greater gas turbine duty than the XtM cases. Therefore, constant feedstock and complete production substitution between urea and electricity is also possible. The

gas turbine load is reduced to 52% when electricity production is substituted by urea production.

IG-PG facilities are very suitable for flexible operation, even though there are technical bottlenecks, significant variation in feedstock and production is possible without drastic reductions in overall efficiencies. The found bottlenecks are:

- To prevent frequent starting-up and shutting down of equipment a minimal 40% load should be maintained in the chemical conversion sections.
- The energy density of biomass is much lower than that of coal. If the gasifier is the limiting factor this results in a significant reduction in energetic input, and thus output, in the facility. This drop is 15% for TOPS and 27% for EP.
- Biomass has a different elemental composition than coal, especially concerning sulphur, chloride and ash content. This can give problems in the gasifier, syngas cooler and Claus unit. Possible solutions are adding flux material, recycling of slag and recycling of sulphur.
- The current maximum co-feeding fraction of biomass is 50% on an energy basis. At these high biomass fractions, the used biomass and coal must be of good quality. This means low moisture content, low ash content and high heating value.
- Producing maximum urea lowers the heating value of the gas stream fed into the gas turbine to below design specification. To maintain a sufficient energy rich gas stream, some of the syngas is directed to the gas turbine instead of the urea synthesis, lowering maximum urea production.
- As the IG-PG facility is designed for multiple chemicals production, optimisation is low. If the IG-PG configuration is limited to only one chemical combined with electricity, optimisation can improve overall efficiency by several %points.

In conclusion, IG-PG facilities have a large potential for flexibility, even with current technology. But, there is a trade-off between flexibility and overall efficiency. Significant variations, both in feedstock and in production are possible with only limited impact on total efficiency. Overall IG-PG facility efficiency drops from 40% to 38% for power; 60% to 55% for FT; 53% to 49% for MeOH and 51% to 47% for urea production when substituting pure coal for pure EP.

When operating the gasifier at maximum capacity, the biggest effect is the reduction in input, and thus output, when switching from coal to biomass. Torrefaction of biomass to TOPS reduces both the drop in plant efficiency and the drop in input and thus output compared to using raw biomass.

The CO_2 emissions are higher than in a dedicated IGCC-CCS facility. This is primarily a direct result of the used assumption that the H_2:CO ratio of the sweet syngas is always equal to the most demanding chemical conversion section, the FT-conversion. Optimisation of the facility could increase CO_2 capture rates to levels comparable of dedicated static IG-PG facilities. But this could decrease flexibility. Especially when only one chemical production is considered, CO_2 capture rates could be improved.

When combining biomass with CCS, CO_2 neutral production is achieved at biomass fraction around 40% on an energy basis. Higher biomass fractions can result in net negative CO_2 emissions.

Topics for further research are the effect of prolonged part-load operation on component availability and efficiency. It should also be noted that the results given in this study are based on computer simulations. As no flexible IG-PG facilities exist, the results cannot be compared with experimental values. Finally, the possibilities and effects of co-feeding large fractions of biomass (up to 100%) have not been tested in EF gasifiers.

Flexibility is probably more important for chemical-electricity substitution than for chemical-chemical substitution. The economics of flexible IG-PG facilities will be investigated in part B.

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Appendix A. Supplementary data

Supplementary data associated with this article can be found, in the online version, at doi:10.1016/j.rser.2011.03.018.

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